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A Subcontract Report

Chem Systems, Inc.
Tarrytown, NY

Prepared under Subcontract No. XX-3-03097-2

Solar Energy Research Institute
A Division of Midwest Research Institute
1617 Cole Boulevard
Golden, Colorado 80401

Operated for the
U.S. Department of Energy
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I. EXECUTIVE SUMMARY

A. Background

Chem Systems has been involved in preliminary design and evaluation of enzyme hydrolysis processes for converting lignocellulose to ethanol, based primarily on its work for SERI in earlier subcontracts. In this earlier work, Chem Systems developed a process design and techno-economic evaluation of a prototype enzyme hydrolysis facility, as well as a computer simulation program which integrated material and utility balances, capital cost estimates and cost of production analyses. A number of parametric studies were performed with the enzyme hydrolysis simulation in an effort to optimize the pretreatment and post hydrolysis options. As a result of these parametric analyses, a process based on enzyme hydrolysis incorporating enzyme production for the conversion of cellulose to hexose sugars seemed to have commercial potential.

Dilute acid prehydrolysis was determined to be one of the most economic pretreatments for enzyme hydrolysis. This concept is the subject of a patent issued to Hans E. Grethlein and assigned to Dartmouth College (U.S. Patent No. 4,237,226, Dec. 2, 1980) entitled "Process for Pretreating Cellulosic Substrates and for Producing Sugar Therefrom." Extensive data covering a wide range of acid prehydrolysis and enzyme hydrolysis conditions were developed by the Thayer School of Engineering at Dartmouth College utilizing the Rutgers C-30 cellulase enzyme. While considerable work is still necessary to optimize the acid prehydrolysis/enzyme hydrolysis configuration and to refine a kinetic model for this configuration, Chem Systems proposed that this concept warranted a detailed feasibility study.

3. Site and Feedstock Selection

One aspect of this feasibility study is the selection of a Midwest site for the ethanol facility and the selection of the wood feedstock based on available forest types. This selection process involved screening ethanol consumption patterns, gasoline demand, applicable tax laws, wood resource acreage and cost elements for delivered wood including both current and future trends.

After a review of the above issues, it was determined that Michigan is probably the most advantageous state within the Upper Midwest and North Central states in which to locate the proposed ethanol plant. The state has the advantages of an industrial base, significant gasoline demand (about 4 percent of national demand and a higher percentage of unleaded gasoline), proximity to other major state markets, adequate wood resources, favorable tax and investment climate, and lack of significant local alcohol production (only 8 million gallons per year), as well as a relatively high state tax exemption for ethanol usage.

The proposed plant would be located in Michigan's Northern lower peninsula, which contains about 1.8 million acres of aspen forest as well as significant quantities of other suitable hardwoods. If it is assumed that 30 percent of the plant feedstock will be hardwoods from aspen forests and the remainder from maple/birch forests, the average feedstock composition will be 57 percent aspen, 20 percent maple and 23 percent other hardwoods. On a moisture-free basis, the plant feed will contain approximately 47.2 percent cellulose, 31.3 percent hemicellulose and 18.5 percent lignin. The hemicellulose can be further broken down to 7.9 percent hexosans, 16.5 percent pentosans and 6.9 percent others.

Gladwin County/Bay City, Midland area has been chosen as the specific location for this facility. This county contains 7 percent of the total state aspen forest land which, by itself, could supply approximately half the required plant feedstock. When coupled with four surrounding counties, the total aspen forest resource base would be sufficient for a facility of twice the capacity of this project. Furthermore, this site provides convenient access to product and by-product markets within the state.

The major by-products from this facility are carbon dioxide and furfural. It is anticipated that the carbon dioxide can be sold locally for use in beverages or enhanced oil recovery. Furfural is currently used primarily in the Midwest and Southwest to produce furfural alcohol, furan resins and THF. Low-cost furfural would open up possibilities of additional THF capacity as well as adipic acid and 1,4-butanediol manufacture. A wide variety of volume chemicals could be produced from these building blocks if the furfural could be priced at approximately half its current market value. Such a scenario is feasible when the furfural is produced as a by-product from an ethanol-from-wood complex.

C. Process Description

The ethanol facility is designed to produce 25 million gallons per year of anhydrous fuel-grade material from mixed hardwood chips, primarily aspen. Major by-products are furfural and carbon dioxide. The plant is divided into eleven processing sections as follows:

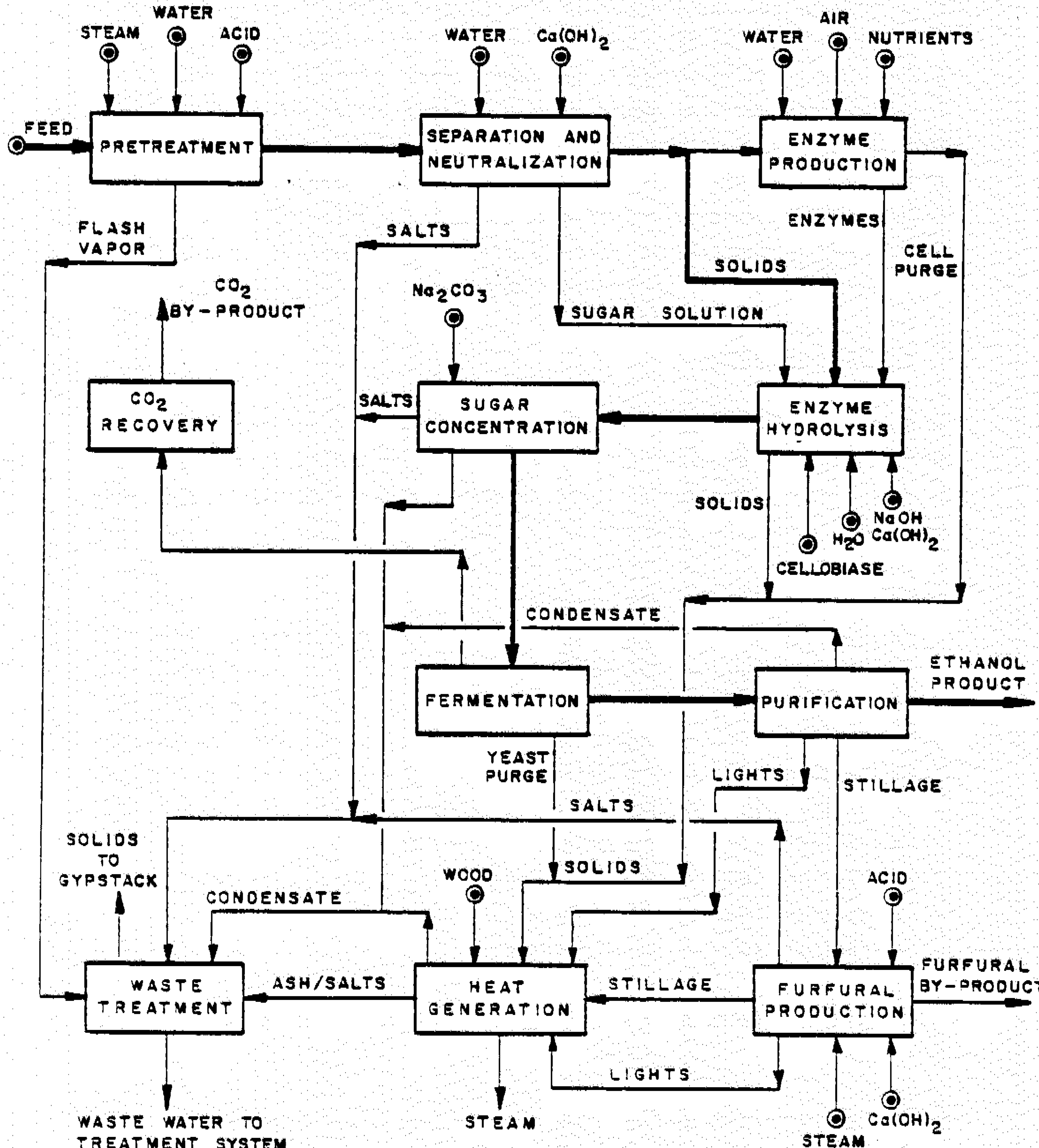
- Section 100 -- Pretreatment
- Section 200 -- Sugar separation and neutralization
- Section 300 -- Enzyme hydrolysis
- Section 400 -- Enzyme production
- Section 500 -- Sugar concentration
- Section 600 -- Fermentation
- Section 700 -- Ethanol purification
- Section 800 -- Carbon dioxide recovery
- Section 900 -- Furfural production
- Section 1000 -- Heat generation
- Section 1100 -- Waste treatment

A block diagram depicting the facility by plant section is shown in Figure I-C-1.

Wood chips are transferred from the storage area via a conveyor to a disk refiner where particle size is reduced sufficiently for compatibility with

FIGURE I-C-1

CELLULOSE TO ETHANOL VIA ENZYME HYDROLYSIS



CHEM SYSTEMS INC.
PROJECT NO. 6042 DATE

prehydrolysis and hydrolysis equipment. This machine requires wood steaming to soften the wood before it is refined into fibers or particles. The steaming of the wood also serves as a preheat for prehydrolysis. Thus, the refiner effluent is fed directly to the prehydrolysis reactor without intermediate pumping. The required acid (0.75 weight percent) and dilution steam/water are mixed with this solids stream to give 15 weight percent solids at the entrance to the plug flow prehydrolysis reactor. Within this reactor the hemicellulose fraction and amorphous cellulose are converted to their respective sugars. The reactor product is flashed in three stages to remove some water/furfural vapor and then proceeds to the centrifuge. Solids from this centrifuge are sent to enzyme hydrolysis and enzyme production. Hot liquids from the centrifuge are sent to a polishing filter to remove the remaining cellulose. The liquid is then neutralized while hot and the precipitated calcium sulfate is filtered out. The liquid stream proceeds to enzyme hydrolysis.

In enzyme hydrolysis, the solids are diluted with water and with the prehydrolysis sugar solution to 10 weight percent and pH adjusted prior to the hydrolyzers. Enzymes from enzyme production are added as well as purchased QM329 cellobiase enzyme (Aspergillus phoenicius). After 24 hours residence time in the cascade reactor system, the yield of glucose on cellulose is 90 percent. The resultant sugar solution contains some unconverted cellulose, which is filtered and washed, removing the bulk of the unconverted cellulose which is pressed and sent to heat generation. The filtrate is pumped to concentration.

Approximately 12 percent of the prehydrolyzed wood is diverted to the enzyme production section where it is used as the carbon source for the RUT C-30 cellulase enzyme (Trichoderma reesei). The quantity of the RUT C-30 required is approximately ten times as large as the QM329 on a volume basis. The RUT C-30 is produced in a fed-batch cycle at a net solids concentration of 15 weight percent taking approximately 12 days. In addition to cellulose, air and nutrients are introduced to the enzyme production tanks. The product from these tanks is sent to a cell centrifuge to remove most of the mycellium, which is then repulped, filtered and washed. Part of the filter cake is recycled to enzyme production to maintain the initial cell concentration in the production tank. The remainder of the filter cake is sent to heat generation. The centrifuge overflow and filtrate are sent to enzyme hydrolysis.

The sugar solution from enzyme hydrolysis is concentrated to 15 percent sugar in a multi-effect evaporator, pH adjusted and sent directly to continuous cascade fermentation. The continuous fermentation step includes yeast separation and recycle with a purge to heat generation. Carbon dioxide produced during fermentation is also recovered for sale. The ethanol product from fermentation is sent to purification, which consists of a beer still and azeotropic dehydration. The ethanol is concentrated to the water-ethanol azeotropic composition in the beer still, then dehydrated by distillation with an azeotroping agent to produce the anhydrous ethanol product. Aqueous stillage from alcohol purification is sent to furfural production.

In the furfural production section, the stillage is heated to approximately 220°C and sulfuric acid is added. After providing sufficent residence time, the reactor product is flashed and neutralized and the precipitated calcium sulfate is filtered out. The flash vapors are cooled by exchange in the furfural-water azeotrope column reboiler, combined with the neutralzied liquids and fed to the azeotrope column. The furfural is further concentrated in the dehydration column. A lights column is utilized to remove low boiling material from the decanted aqueous layer originating from the azeotrope column overheads.

Hydrolysis solids purge, furfural production stillage, waste yeast, lights and waste enzymes are sent to a multi-effect evaporator prior to being used as fuel in the steam boiler where 1,200 psia steam is raised. Additional wood chips are fed to this steam boiler to bring the plant steam requirements into balance. Ash from the boiler and salts from neutralization are sent to a gypstack.

The condensate from various process concentration steps is combined with dilute aqueous waste streams and sent to a waste treatment system. Treated water is reused in the process.

The offsites included in the design are storage facilities, a cooling water system, a wood-fired steam boiler with electricity cogeneration facilities, boiler feedwater pretreatment, buildings, piping, pollution control, general utilities and land site development.

D. Plant Economics

An economic evaluation has been performed for the wood-based ethanol plant described in the preceding section. The capital cost for a 25 million gallon per year plant is estimated at 112.0 million dollars based on mid-1984 costs at a Michigan site location. This estimate includes 66.3 million dollars for inside battery limits cost (ISBL) and 45.7 million dollars for offsites. Summaries of the ISBL and offsites costs are presented in Tables I-D-1 and I-D-2, respectively.

A 1984 manufacturing cost summary for the 25 million gallons per year ethanol plant, shown in Table I-D-3, serves as the basis for a cash flow analysis over the project life.

The raw material costs is based on a current-hardwood price of 18 dollars per wet ton which is determined based on an average hauling distance of 50 miles.

The total cash cost of production for ethanol is estimated at 36 cents per gallon. The net cost of production for ethanol, including depreciation, is 157 cents per gallon and the instantaneous mid-1984 ethanol sales price is 206 cents per gallon with a 10 percent DCF return on investment.

TABLE I-D-1

ISBL* CAPITAL COST SUMMARY, MID-1984

Section	Name	Total Field Installed Cost, $
100	Pretreatment/prehyrolysis	3,104,000
200	Sugar separation and neutralization	3,901,000
300	Enzyme production	6,376,000
400	Enzyme hydrolysis	7,069,000
500	Sugar concentration	5,941,000
600	Fermentation	4,209,000
800	Ethanol purification	4,111,000
900	Furfural production	5,166,000
1000	Heat generation	3,311,000
Total field cost		43,188,000
Engineering		4,180,000
Construction overhead		4,850,000
Bare plant cost		52,218,000
Contingency		5,222,000
Contractors fee		1,509,000
Special charges		1,550,000
Total facility cost		60,490,000
CO_2 recovery system		5,800,000
Total ISBL		66,290,000

*Section 1100, Waste treatment, is included in offsites costs.

TABLE I-D-2

OFFSITES CAPITAL COSTS, MID-1984

	Total Installed Cost, $
Total storage	1,928,000
Steam boiler	11,700,000
Generator/turbine/switch gear	6,600,000
Wood handling system	8,100,000
Cooling water	2,760,000
Electrical system	982,000
Boiler feedwater system	490,000
Pollution control	
Pond and gypstack (Section 1100)	2,560,000
Sanitary waste	300,000
Buildings	1,815,000
General utilities	3,024,000
Site development	3,530,000
Piping	1,815,000
Total offsites	45,704,000

TABLE I-D-3

ETHANOL MANUFACTURING COST SUMMARY
Mid-1984, Michigan Site

	Cents per Gallon
Raw materials(1)	80.97
By-product credit(2)	(62.04)
Utilities.	19.89
Operating costs	22.65
Overhead expenses	24.47
Cash cost of production	85.95
Depreciation	71.33
Net cost of production	157.28
Ethanol sales price at 10 percent DCF return	206.0

(1) Wood feedstock at $18 per wet ton.
(2) Furfural and carbon dioxide by-product values at 30 and 2.8 cents per pound, respectively.

Using the 1984 cost data as the basis, a cash flow analysis was performed to determine the internal rate of return (IRR) for the project. Table I-D-4 highlights the economic parameters used in this analysis.

TABLE I-D-4

ECONOMIC PARAMETERS FOR THE CASH FLOW ANALYSIS

Plant operation (after start-up): 1987 through 2001
Construction period: 3 years
Capacity build-up: 60 percent first year
80 percent second year
100 percent remaining years
Interest rates (for debt financing): 13 percent
MSARD: Six percent of revenues
Depreciation: Straight line method over five years for ISBL capital and ten years for offsites
Investment tax credit: 10 percent of total capital credited against current or future tax liabilities
Corporate tax: 45 percent of profit
Federal/state tax incentive: 9 cents per gallon for 10 percent ethanol blends in gasoline

Since the cash flow is based on current dollars the cost elements were escalated over the project life. These escalation rates are presented in Table I-D-5. Revenues are generated from ethanol along with by-product furfural and carbon dioxide. The ethanol market value over the life of the project is forecast based on the regional demand for gasoline by year, grade and distance from the Gladwin County plant site. The federal tax incentive for 10 percent ethanol blends in gasoline was projected at six cents per gallon and the state tax incentive was assumed to be three cents per gallon for the project life. Combining these factors, ethanol values at the plant gate were estimated for the 1987-2001 time period. Table I-D-6 summarizes this ethanol price forecast.

TABLE I-D-5

ESCALATION RATES, 1984-2001
(Percent per Year)

Period	GNP Deflator (1)	Capital Cost (2)	Utility Costs
1984-1990	5.0	6.5	7.5
1990-2001	4.5	5.5	6.9

(1)Used for feedstock, labor and direct overhead costs.
(2)Used for capital, maintenance, general plant overhead and insurance.

TABLE I-D-6

ETHANOL PRICE FORECAST

Year	Price, Cents/Gallon
1987	200.8
1988	207.4
1989	214.4
1990	221.5
1991	227.7
1992	234.2
1993	240.7
1994	247.5
1995	254.5
1996	261.9
1997	269.5
1998	277.3
1999	285.3
2000	293.6
2001	302.1

For a plant that is 100 percent equity financed, the IRR for this project is 5.9 percent. If the plant is based on 30 percent equity and 70 percent debt financing the IRR becomes negative since the after tax cost of money is greater than the equity return.

E. Conclusions and Recommendations

The enzyme hydrolysis based ethanol facility evaluated in this study is a feasible project from both technical and economic points of view. The strengths and weaknesses of this project are discussed in this section.

The technology base for this facility is relatively sound. The major area where further research is necessary involves enzyme production using prehydrolyzed wood. While this has been done, the yields and process conditions assumed in this study have not yet been attained. Reasonable conditions have been assumed based on experimental evidence and it is expected that research with pretreated wood will reveal methods of handling that equal or even exceed the expectations of this study.

The remainder of the process steps through enzyme hydrolysis itself have been demonstrated individually on a small scale. It remains to verify that these steps can be carried out in an integrated fashion in industrial-type equipment. This is especially true of prehydrolysis and the various solid/liquid separations encountered. Integration of process operations will reveal if there are any inhibitory effects from minor constituents formed in pretreatment and, if so, how to deal with them.

The rest of the process, from glucose fermentation through furfural production, is based on commercially proven technology. Other than verifying that sugars from enzyme hydrolysis are fermentable, no further development of this half of the process is necessary.

In addition to 25 million gallons per year of fuel-grade ethanol, this facility coproduces 160 million pounds per year of liquefied carbon dioxide and 36 million pounds per year of furfural. Revenues from these by-products help to offset the effects of the substantial capital investment (112 million dollars in mid-1984). Furthermore, virtually all waste organics are utilized to provide steam. Additional heat is provided by additional wood chips to bring the plant into thermal balance. More costly fuels are avoided. Most of the plant electricity requirements are met by cogeneration from plant steam. Thus, utility costs are minimal.

Evaluating this facility on a mid-1984 basis with wood chips at 18 dollars per wet ton and by-product furfural at 30 cents per pound results in a required ethanol sales price of 206 cents per gallon at 10 percent DCF return on investment. On the same economic basis a 50 million gallon per year ethanol plant based on whole corn kernel milling would yield a sales price of 219 cents per gallon. Thus, the innovative enzyme hydrolysis based facility should be more than competitive with current commercial fermentation technology.

The choice of Gladwin County in Michigan for the plant site results in an ethanol market almost entirely within a 200-mile radius. Furthermore, there are adequate hardwood feedstock supplies within a 50-mile radius of the plant site. An ethanol pricing scenario incorporating a combined federal and state tax incentive of nine cents per gallon for 10 percent ethanol blends in gasoline was developed for Michigan over the project life and then used to generate the project cash flow analysis. This resulted in an internal rate of return of 5.9 percent based on 100 percent equity financing. Although profitable, this return is probably not high enough to convince investors to go ahead with such a venture at this time. However, optimization studies have pointed out reasonable targets for improvement in enzyme production and enzyme hydrolysis. If some of these improvements could be achieved through additional research and development, the proposed ethanol facility would become commercially interesting.

In conclusion, it seems that an enzymatic hydrolysis based ethanol plant using prehydrolysis pretreatment is a viable process concept. Research to optimize the enzyme production and hydrolysis sections of the facility is still required and then the front half of the process must be demonstrated in industrial-type equipment to provide proof of concept. When these steps are successfully completed, the process would be ready for commercialization. A central Michigan location would be ideal for the first plant of this kind.

II. INTRODUCTION

Chem Systems has been involved in preliminary design and evaluation of enzyme hydrolysis processes for converting lignocellulose to ethanol, based primarily on its work for SERI in earlier subcontracts. In this earlier work, Chem Systems developed a process design and techno-economic evaluation of a prototype enzyme hydrolysis facility. Using the design as a basis, Chem Systems developed a computer simulation program which integrated material and utility balances, capital cost estimates and cost of production analyses. The user was able to choose pretreatement options and most of the operating parameters. A number of parametric studies were performed with the enzyme hydrolysis simulation in an effort to optimize the pretreatment and post hydrolysis options.

As a result of parametric analyses using the simulation program, a process based on enzyme hydrolysis incorporating enzyme production for the conversion of cellulose to hexose sugars seemed to have commercial potential. There are many potentially interesting process configurations incorporating enzyme hydrolysis. However, the most viable configurations seem to be those which involve minimial pretreatment steps and still maximize the recovery of useable by-products.

Dilute acid prehydrolysis has been determined by Chem Systems to be one of the most economic pretreatments for enzyme hydrolysis. This concept is the subject of a patent issued to Hans E. Grethlein and assigned to Dartmouth College (U.S. Patent No. 4,237,226, Dec. 2, 1980) entitled "Process for Pretreating Cellulosic Substrates and for Producing Sugar Therefrom." Extensive data covering a wide range of acid prehydrolysis and enzyme hydrolysis conditions were correlated and incorporated into the simulation. These data were developed by the Thayer School of Engineering at Dartmouth College utilizing the Rutgers C-30 cellulase enzyme. Further parametric studies with the model verified that acid prehydrolysis is indeed an economic pretreatment for enzyme hydrolysis since a portion of the hexosan is converted to sugar in this step and the remaining cellulose becomes very accessible to enzyme attack.

While considerable work is still necessary to optimize the acid prehydrolysis/enzyme hydrolysis configuration and to refine the kinetic model for this configuration, Chem Systems proposed that this concept warranted a detailed feasibility study.

III. PROCESS CHEMISTRY

A. Wood Chemistry

Wood consists chiefly of hollow interconnected fibers axially oriented in the tree. These support the tree and form the conduits for the transport of water from the roots to the leaves where, under the catalytic action of chlorophyll, the water reacts with carbon dioxide to form sugars and other organic materials. Growth of the tree results from the return of the solution of these substances through the inner bark to further react and polymerize at the cambium at the interface between wood and bark, thus forming new cells which in turn form a further ring in the fibrous conduit system.

The conifers, of which spruce is an example, are generally referred to as softwoods. The broad-leaved deciduous woods which are morphologically and chemically distinct from the conifers are commonly referred to as hardwoods in spite of the fact that certain species, such as basswood and poplar, have woods which are relatively soft.

The structural components of wood, which comprise some 90 percent of the weight of the wood, consist of the three natural polymers cellulose, lignin, and hemicelluloses. The 3 to 10 percent nonstructural components of wood, such as terpenes and waxes, are referred to as extractives.

Cellulose is a highly oriented, crystalline, linear polymer of glucose units. Values for the degree of polymerization of cellulose chains in wood range from 7,000 to 10,000. Cellulose can be hydrolyzed to glucose by acids whereupon the glucose formed by cellulose hydrolysis can be fermented to ethanol under the influence of enzymes from yeast.

Lignin is a 3-dimensional random polymer formed from phenylpropane units. It acts as a cement between the cellulose fibers and as a stiffening agent within the fibers. Its molecular weight is over 10,000.

Hemicelluloses are polysaccharides which are short or branched polymers of 5-carbon sugars such as xylose or 6-carbon sugars other than glucose. Hardwoods contain about 30 percent hemicelluloses, of which the principal constituent sugars in decreasing abundance are xylose, galactose, and mannose. Degrees of polymerization range from less than 100 to about 200 sugar units. Hemicelluloses are readily hydrolyzable to xylose and other simple sugars by acids. Mannose and other hexoses can be fermented to ethanol, while xylose and other pentoses are readily dehydrated to furfural by acids.

The wood cell wall is then a fiber-reinforced plastic with cellulose fibers embedded in an amorphous matrix of hemicelluloses and lignin.

The structure and composition of bark are very different from that of wood. Typically, the cellulose content of bark is only 20 to 30 percent. Bark contains a significant proportion of extractives, generally 20 to 40 percent. Extractive-free bark contains two components not found in wood, suberin and phenolic acids, as well as cellulose, hemicellulose, and lignin. Suberin is a hydroxy acid complex consisting of esters of higher aliphatic hydroxy acids and phenolic acids. Phenolic acids are high molecular weight phenols that differ from lignin in their lower molecular weight, high carboxyl content, and lower methoxyl content.

In hardwoods, the principal hemicellulose is 4-O-methylglucurono xylan acetate. This is made up of 1 part 4-O-methyl-alpha-D glucuronic acid, 10 parts beta-D xylose, and 7 parts O-acetyl-beta-D-xylose.

Cellulose is a long chain polymer of beta-D-glucose in the pyranose form, linked together by 1,4-glycosidic bonds to form cellobiose residues that are the repeating units in the cellulose chaining. The beta-linkage requires that the alternate glucose units must be rotated through 180 degrees. An important implication of this structure is a marked tendency for the individual cellulose chains to come together to form bundles, of a crystalline nature, held together by hydrogen bonds.

Lignins are 3-dimensional network polymers formed from phenylpropane units that have randomly grown into a complicated large molecule with many different kinds of linkages between the monomers. Although hardwood lignins differ somewhat in composition from softwood lignins, chiefly in methoxyl substitution and the degree of carbon-carbon linkage between phenyl groups, the common structural features predominate. This includes an aromatic and phenolic character as well as covalent carbon-carbon bonding that prevents reversion to monomers by mild processing. The random structure arises from an enzymatically initiated free radical polymerization of lignin precursors in the form of p-hydroxy cinnamyl alcohols. In conifers, the precursor is principally coniferyl alcohol (3-methoxy-4-hydroxycinnamyl alcohol). Coniferyl alcohol yields the so-called quaiacyl lignin. In hardwoods, additional precursors such as sinapyl alcohol (3,5-dimethoxy-4-hydroxycinnamyl alcohol) and coumaryl alcohol (4-hydroxycinnamyl alcohol) are also present resulting in the so-called quaiacyl-syringyl lignins. Because the quaiacyl unit has an additional potential reactive site instead of the extra methoxyl group in the syringyl unit, a higher degree of cross-linking exists in quaiacyl lignins. The lower apparent molecular weight and easier dissolution of hardwood lignins are manifestations of their lower degree of cross-linking. The small quantities of methanol produced as a by-product of the wood hydrolysis process arise from the methoxyl groups ($-OCH_3$) in lignin.

The specific composition of wood varies from species to species. Table III-A-1 shows the typical composition of several woods. It can be seen from the table that, in general, hardwoods have a greater holocellulose (cellulose and hemicellulose) content than softwoods, although the hexosan content is approximately equivalent for both, due to a lower percentage of xylans in the softwood hemicellulose.

Table III-A-2 indicates the potential reducing sugars that can be produced from hardwoods and softwoods. It shows that aspen has the greatest potential for producing reducing sugars which can subsequently be fermented. Populus Tremuloides (poplar) or trembling aspen has high availability in North America and can be grown as a high yield, short

TABLE III-A-1

PERCENTAGE COMPOSITION OF CERTAIN WOODS
(Extractive Free, Ash Free)

	Lignin	Holo-cellulose	Alpha-cellulose	Hemi-cellulose	Pentosans	Uronic Acid Anhydride	Acetyl	Methoxyl in Carbohydrate
Softwoods								
White Spruce	26.6	73.3	49.5	23.8	10.9	2.68	2.35	0.70
Red Spruce	26.6	72.9	48.3	24.6	11.6	3.20	2.50	0.92
Eastern Hemlock	31.5	68.5	48.2	20.3	10.0	3.40	1.87	0.81
Balsam Fir	30.1	69.9	44.0	25.9	10.3	3.08	2.24	0.41
Jack Pine	27.2	72.5	49.5	23.0	12.8	2.92	1.92	0.75
Hardwoods								
Aspen	17.3	82.5	50.7	31.8	23.5	4.28	4.65	0.93
Willow	22.0	78.3						
Maple	23.5	76.3	50.0	26.3				
White Oak	24.1	75.4	49.5	25.9				

rotation crop in many parts of the northern United States. These reasons make aspen an excellent major feedstock for the enzymatic hydrolysis-based ethanol process described herein.

TABLE III-A-2

YIELD OF POTENTIAL REDUCING SUGARS AND FERMENTABLE SUGARS FROM SAMPLES OF REPRESENTATIVE HARDWOODS AND SOFTWOODS

	Potential Reducing Sugars, %	Fermentability, %	Potential Fermentable Sugars, %
Hardwoods			
American Beech	70.1	75.1	52.6
Aspen	75.1	76.3	57.3
Birch	69.9	67.8	47.4
Maple	68.2	71.0	48.4
Red Oak	63.6	63.0	40.2
Sweetgum	66.4	73.8	49.0
Yellow Poplar	70.9	76.1	54.0
Softwoods			
Douglas Fir	66.6	86.2	57.4
Eastern White Pine	66.5	86.3	57.4
Hemlock	66.1	88.2	58.3
Ponderosa Pine	68.0	82.2	55.9
Redwood	52.4	77.1	40.4
Sitka Spruce	70.1	85.3	59.8
Southern Yellow Pine	64.8	82.0	53.2
Sugar Pine	64.8	82.4	53.0

B. Pretreatment

Two methods of pretreating wood feedstock, steam explosion and dilute acid prehydrolysis, have been proposed recently as efficient methods of rendering cellulose accessible to enzymatic attack thus resulting in high glucose yields. The latter pretreatment, based on the developmental work of the Thayer School of Engineering at Dartmouth College(1), is the basis for this study.

Dilute acid prehydrolysis is a pretreatment designed primarily to attack and convert the hemicellulose and amorphous cellulose fractions of wood.

Because of the relatively low reaction temperature, only a small fraction of the crystalline cellulose is converted to glucose. Furthermore, the amount of degradation products formed from the resulting hexose and pentose sugars is minimal. Thus, the net result of prehydrolysis is the formation of soluble sugars from the hemicellulose fraction which leaves a residual higher porosity solid material that is more susceptable to enzymatic attack.

A plug-flow reactor kinetic model has been developed at Dartmouth for this system[2,3]. These kinetics assume that the acid-catalyzed hydrolysis of crystalline cellulose to glucose can be described as a homogeneous pseudo-first-order reaction and that the acid-catalyzed decomposition of glucose is also a first-order reaction. Amorphous cellulose and hemicellulosic hexosans hydrolyze immediately to glucose and can be taken as an initial glucose fraction. Thus, the hydrolysis can be described by:

$$\text{Crystalline Cellulose} \xrightarrow{K_1} \text{Glucose} \xrightarrow{K_2} \text{Decomposition Products}$$
$$\text{Amorphous Cellulose + Hemicellulose} \longrightarrow \text{Glucose}$$

The integrated kinetic expressions describing these reactions are:

$$G(t) = C(o) \left[\frac{K_1}{K_1 - K_2} \right] \left[\exp(-K_2 t) - \exp(-K_1 t) \right] + G(o) \exp(-K_1 t) \qquad (1)$$

$$C(t) = C(o) \exp(-K_1 t) \qquad (2)$$

where

$$K_i = P_i A^{Ni} \exp(-E_i/RT)$$

C is the fraction of potential glucan remaining as crystalline cellulose, G is the fraction of potential glucan present as glucose and A is the weight percent sulfuric acid in the aqueous phase. The cellulose decomposition product is assumed to be hydroxymethyl furfural although further reaction will result in some levulinic acid and formic acid. The activation energies, exponents and pre-exponential factors are tabulated for different wood feedstocks in Table III-B-1. It can be seen that some variation in cellulose decomposition rate is expected for different wood feedstocks.

TABLE III-B-1

CELLULOSE DECOMPOSITION KINETIC CONSTANTS

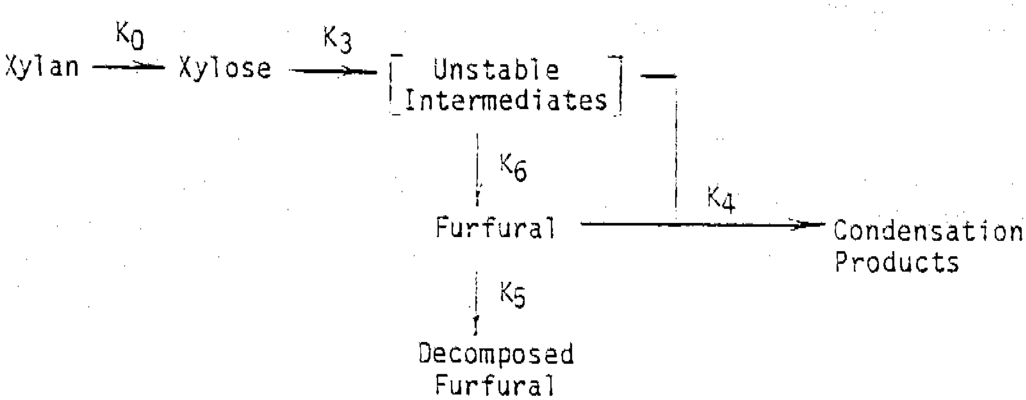
GLUCOSE FORMATION[2,3]

	P_1, min^{-1}	N_2	E_1, cal/g mol
Poplar	6.12×10^{15}	0.99	35,150
White Pine	7.80×10^{13}	0.96	30,170
Mixed hardwoods (90% birch, 10% maple)	8.98×10^{20}	1.55	47,100

GLUCOSE DECOMPOSITION[3]

	P_2, min^{-1}	N_2	E_2, cal/g mol
Glucose	3.96×10^{8}	0.57	21,000

Decomposition kinetics of the hemicellulosic xylan have also been investigated in a plug-flow reactor system at Dartmouth[4]. This hydrolysis sequence can be described by:



The rate equations describing these reactions are:

$$\frac{dH}{dt} = -K_0H \tag{3}$$

$$\frac{dX}{dt} = K_0H - K_3 X \tag{4}$$

$$\frac{dD}{dt} = -K_5F \tag{5}$$

$$U = 0; \frac{dU}{dt} = 0 \tag{6}$$

$$\frac{dF}{dt} = K_3X - K_5F - K_7XFQ_0 \tag{7}$$

where

$$K_7 = K_3K_4/K_6$$

and

- H = Mole fraction of potential xylose as xylan.
- X = Mole fraction of potential xylose as xylose.
- U = Mole fraction of potential xylose as unstable intermediates.
- F = Mole fraction of potential xylose as furfural.
- D = Mole fraction of potential xylose as decomposed furfural.
- C = Mole fraction of potential xylose as condensation products.
- Q_0 = Concentration in moles/liter of potential xylose in feed.

The activation energies, exponents and pre-exponential factors for xylan decomposition in mixed hardwood are shown in Table III-B-2.

TABLE III-B-2

KINETIC PARAMETERS FOR MIXED HARDWOOD XYLAN DECOMPOSITION[4]

	K_0	K_3	K_5	K_7
E (cal/gmol)	27,826	27,130	15,279	23,943
Ln P (min^{-1})	31.76	28.48	13.23	26.74
N	1.17	0.688	0.579	0.517

Solving these equations for an aspen/mixed hardwood feed to a 12-second plug flow reactor at 200°C with 0.75 weight percent sulfuric acid results in the conversion of 5 percent of the crystalline cellulose. Since all the hemicellulose hexosans are converted, the total C_6 hydrolyzed is 13.6 mole percent with a selectivity to glucose of 98.5 percent. On the other hand, approximately 71 percent of the xylan is converted with a selectivity to xylose of 93.5 percent and a selectivity to furfural of 6.4 percent. Thus, these conditions result in 27 weight percent of the cellulose and hemicellulose converted, primarily to sugars with only minor amounts of further degradation products.

In order to feed the wood to the pressurized prehydrolysis reactor, the chips must be reduced in size to aproximately 1 millimeter. This can be accomplished in a disk refiner which operates most efficiently when the wood chips are steam presoaked. Thus, a digester with a 1-5 minute residence time preceeds the disk refiner. The chip steaming treatment produces acetic acid by the cleavage of acetyl groups in the wood, thereby improving the digestibility of the wood due to the hydrolytic effect of the acetic acid. The effectiveness of that treatment varies with the ash content of the wood, which has a buffering effect on the acid. It has been assumed that 25 percent of the available hemicellulose acetyl groups are converted to acetic acid in 1 minute at 200°C.

Cleavage of methoxy groups in the lignin fraction occurs at these conditions. Based on measurements made at Dartmouth[5], a 0.05 weight percent methanol yield based on dry wood is taken for the prehydrolysis product. Also, some of the lignin is rendered soluble by steam treatment and prehydrolysis. This has been estimated at 10 percent.

C. Enzyme Hydrolysis

Recent advances in enzyme technology, specifically in enzyme productivity, have made enzyme hydrolysis an attractive route for cellulose conversion. Steam pretreatment of cellulose substrates and mild acid prehydrolysis have enabled glucose yields via enzyme hydrolysis to approach theoretical in a relatively short residence time of around 24 hours. Enzyme production costs, once prohibitively high, have been markedly decreased recently by significant increases in enzyme productivity utilizing new mutant strains of the fungus, Trichoderma reesei.

Genetic researchers at Rutgers University have developed a mutation of T. reesei, RUT C-30, which has significantly higher enzyme productivity than the previously used strain, QM9414. In addition, experimentation with various types of nutrient media for enzyme production has enabled cheaper ingredients, such as corn steep liquor and steam exploded wood, to be substituted for the more expensive ingredients previously used such as proteose peptone and delignified cellulose.

The mechanism of enzyme hydrolysis by T. reesei enzymes occurs when crystalline cellulose is attacked by the enzymes endo-beta-glucanase and cellobiohydrolase forming oligosaccharides, including cellobiose. The cellobiose is then acted upon by endo-beta-glucanase and beta-glucosidase to form glucose. This is represented as:

crystalline cellulose ← endo-beta-glucanase

↓

cellobiohydrolase → amorphous cellulose

↓

beta-glucosidase (cellobiase) → cellobiose oligosaccharides ← (endo-beta-glucanase)

↓

glucose

Therefore the enzyme mixture necessary to obtain optimum saccharification is a complex mixture of several enzymes in various proportions. RUT-C-30 has high enzyme productivity, but produces essentially all endo-beta-glucanase with little beta-glucosidase. Natick labs[6] has experimented with various mutant strains and has developed QM329 from Aspergillus phoenicis, which produces beta-glucosidase at very high productivities. Ideally, the optimum enzyme mixture would be a combination of endo-beta-glucanase from RUT C-30 and beta-glucosidase from QM329 in a ratio of cellulase:cellobiase of about 15:1.

Enzyme hydrolysis following dilute acid prehydrolysis pretreatment has been thoroughly investigated at Dartmouth[7]. Based on the prehydrolysis conditions described earlier, a glucose yield of 90 percent of theoretical can be achieved in 24 hours at 50°C with a pH of 4.8. The initial solids concentration can be 10 weight percent. A cellulase loading of 15 international units per gram of solids is required while the cellobiase loading is 0.9. It is assumed that all the xylan present is converted to xylose.

D. Fermentation

The fermentation process involves the conversion of simple hexose sugars to ethanol and carbon dioxide. The many possible hexose sugars are distinguished from one another by subtle structural differences. Relatively few hexose sugars occur naturally in large quantities, but of these the most important for fermentation are glucose, fructose, mannose and galactose.

A great deal of research has been devoted to investigating the metabolism of sugar by yeast. As a result, the complex transformations that take place and the enzymes responsible for these changes are now better understood than for any other organism. The yeast fermentation reactions and their interrelationships are usually designated by the Embden-Meyerhof-Parnas scheme, which shows the importance of organic phosphates (Figure III-D-1).

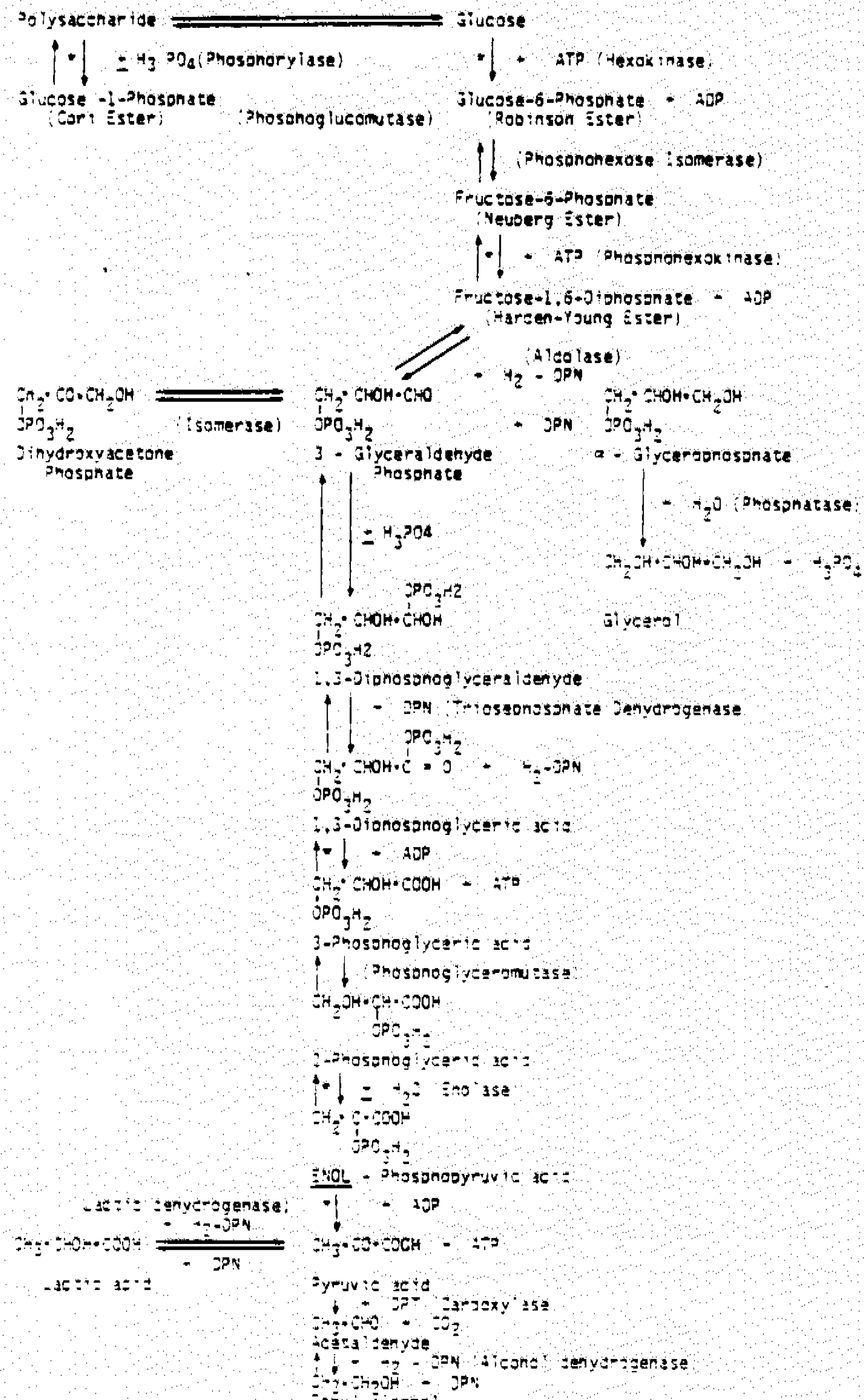
Most industrially important microorganisms metabolize carbohydrates by means of the EMP pathway to pyruvic acid. The overall reaction may be represented by the following equation:

$$C_6H_{12}O_6 + 2\ DPN + 2\ ADP + 2\ P_i \longrightarrow 2\ CH_3CO\ COOH + 2\ H_2DPN + 2\ ATP$$

Since both DPN and ADP are present in only small quantities in the microbial cell, they must be regenerated if metabolism is to continue. The terminal phosphate bond of the ATP is used by energy-requiring reactions in the cell, thus liberating ADP for further phosphorylation. Alternatively, it can by hydrolyzed by ATPase to ADP and inorganic phosphate.

FIGURE III-D-1

EMBDEN - MEYERHOF - PARNAS PATHWAY



*MG— or MN— ions required
ATP Adenosine triphosphate
ADP Adenosine diphosphate
H3PO4 Inorganic phosphate
DPN Diphosphopyridine nucleotide (cozymase)
H2 DPN Reduced diphosphopyridine nucleotide
DPT — Diphosphothiamine (cocarboxylase)

The pyruvic acid formed by the EMP pathway is converted to ethanol and carbon dioxide by the following reactions:

- The enzyme carboxylase catalyzes a decarboxylation of pyruvic acid to acetaldehyde:

$$CH_3COCOOH \xrightarrow[\text{decarboxylase}]{\text{Pyruvate}} CH_3CHO + CO_2$$

  This enzyme requires cocarboxylase DPT as a coenzyme.

- The enzyme alcohol dehydrogenase, working in reverse, catalyzes a reduction of acetaldehyde to ethanol at the expense of reduced DPN:

$$CH_3CHO + H_2DPN \xrightarrow[\text{Alcohol dehydrogenase}]{} CH_3CH_2OH + DPN$$

The fermentation reaction is generally catalyzed by enzymes produced by special strains of yeast developed for efficient fermentation. Although there is much research at present on the production of ethanol using fungi and bacteria, all commercial grain ethanol fermentation processes are presently based on yeast. The most commonly used are strains of Saccharomyces cerevisiae. The criteria for good distillery yeasts are high alcohol and sugar tolerance, efficient conversion at temperatures of at least 32°C, and general hardiness to fairly extreme changes in conditions such as pH, temperature, and osmotic pressure. Fortunately, a relatively high percentage of yeast strains meets these criteria.

In a continuous cascade system with yeast recycle and an initial yeast concentration of 385×10^6 cells per milliliter it is assumed that all the glucose is converted. This requires an 18-hour residence time at 30°C with an initial glucose concentration of 13.2 weight percent. Adjusting the initial pH to 4.0 will result in the following selectivity structure:

- Conversion of 93.0 mole percent to ethanol and carbon dioxide via:

$$C_6H_{12}O_6 \longrightarrow 2C_2H_5OH + 2CO_2$$

- Conversion of 4.9 mole percent to glycerol, acetaldehyde and carbon dioxide via:

$$C_6H_{12}O_6 \longrightarrow C_3H_8O_3 + C_2H_4O + CO_2$$

- Conversion of 0.1 mole percent to fusel oils.

- Remaining 2.0 mole percent utilized for yeast growth with equal weight of carbon dioxide evolution.

The net result is approximately 47.5 pounds of ethanol produced and 47.6 pounds of carbon dioxide produced per 100 pounds of glucose.

E. Furfural Production

The formation of furfural is accomplished by dilute acid hydrolysis of xylose. As explained in Section III-B, this reaction is accompanied by condensation products such as resinous tars and decomposition products such as formic acid. The overall kinetics are dependent on the concentration of potential xylose in the feed with higher furfural yields obtainable at lower xylose concentrations.

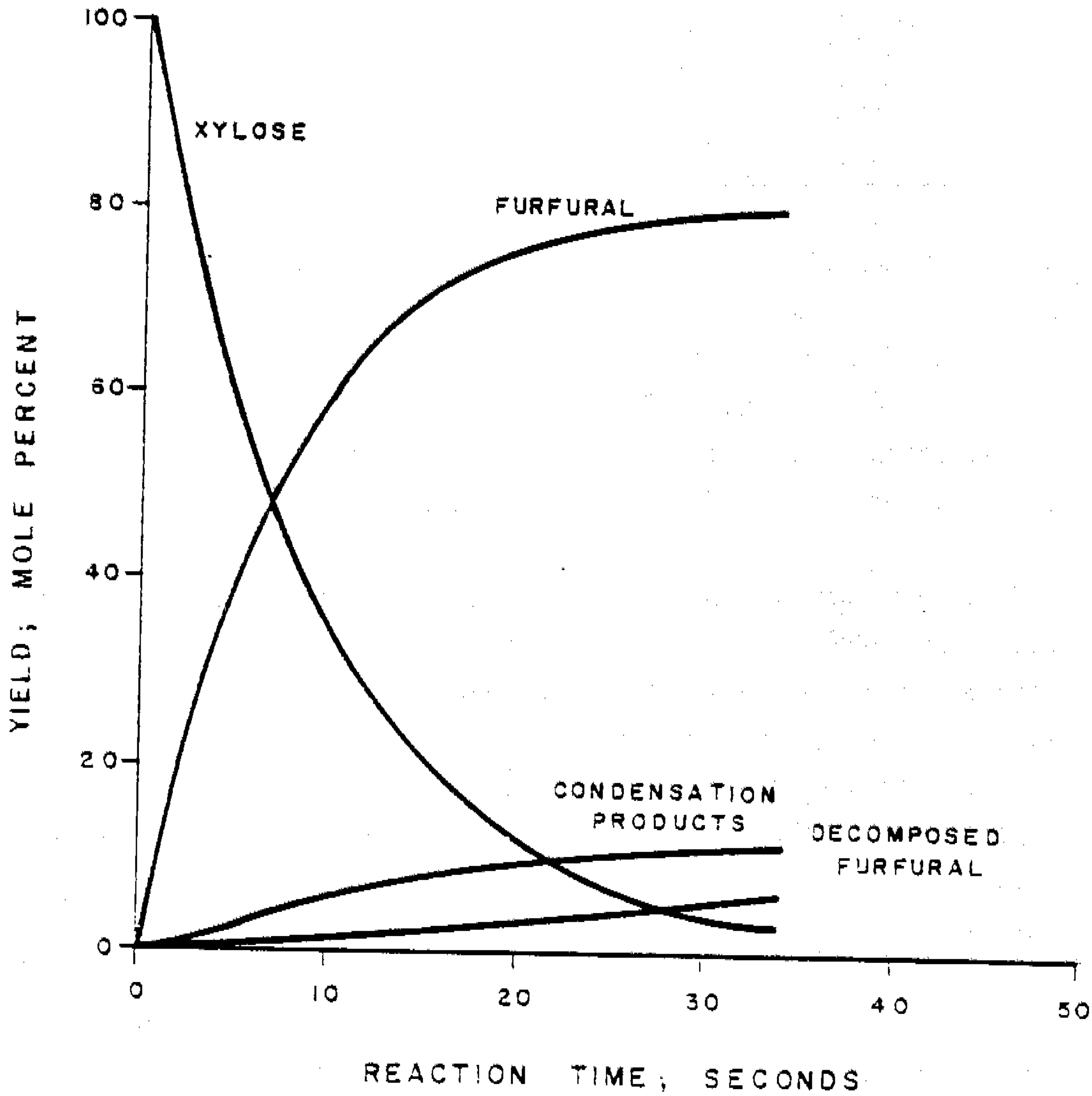
Furfural production takes place in the liquid phase at temperatures in excess of 200°C with sulfuric acid concentrations of 0.5-1.5 weight percent. The residence time can vary from 10 seconds to 10 minutes, depending on reactor temperature. The effect of residence time on furfural and by-product yields is shown in Figure III-E-1 for a 1 weight percent xylose solution; based on equations 3-7 in Section III-B. At this xylose concentration, furfural yields of nearly 80 mole percent are possible.

A more concentrated xylose stream of nearly 5 weight percent, such as encountered in the beer still bottoms in the current design, will result in a 47.6 mole percent furfural yield. This can be accomplished with a

FIGURE III-E-1

XYLOSE DECOMPOSITION AT 240°C WITH 1% ACID

1% XYLOSE



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70-second residence time at 220^{o}C and a sulfuric acid concentration of 1 weight percent. Obviously, the furfural yield could be increased by dilution, but the plant would be penalized by increased steam requirements.

IV. PROCESS DESIGN

A. Process Description

This section provides a process description of a plant designed to produce 25 million gallons per year of anhydrous ethanol from mixed hardwood. Figures IV-1 through IV-11 are process flowsheets for the plant and should be referred to with the following process descriptions. The detailed material balance is located in Section X-A and the design basis for the major processing units is detailed in Section IV-B.

1. Pretreatment/Prehydrolysis (Section 100) (Figure IV-A-1)

The pretreatment/prehydrolysis section can be subdivided into four subsections:

- Chip washing
- Disk refining
- Prehydrolysis
- Reactor effluent flashing

Mixed hardwood chips are transferred from the storage area via a conveyor to a chip washer, VS-101, where any entrained dirt accompanying the wood chips is removed. Untreated water from the waste pond is used for this purpose and the effluent is returned to the waste pond. The washed wood chips are transferred to chip bin VT-101 via a conveyor (Stream 101).

Wood chips are fed from the chip bin to digester CO-104 where they are mixed with 17 atm steam (Stream 102). The digester operates at 16 atm and 200°C with a residence time of one minute. This softens the wood and provides a pressure head for further processing. Thus, the softened wood chips (Stream 103) are discharged into single disk refiner CR-101 which reduces the particles to nominal 1 millimeter diameter short fibers. After size reduction, the wood particles are discharged through a ball valve into the prehydrolysis reactor.

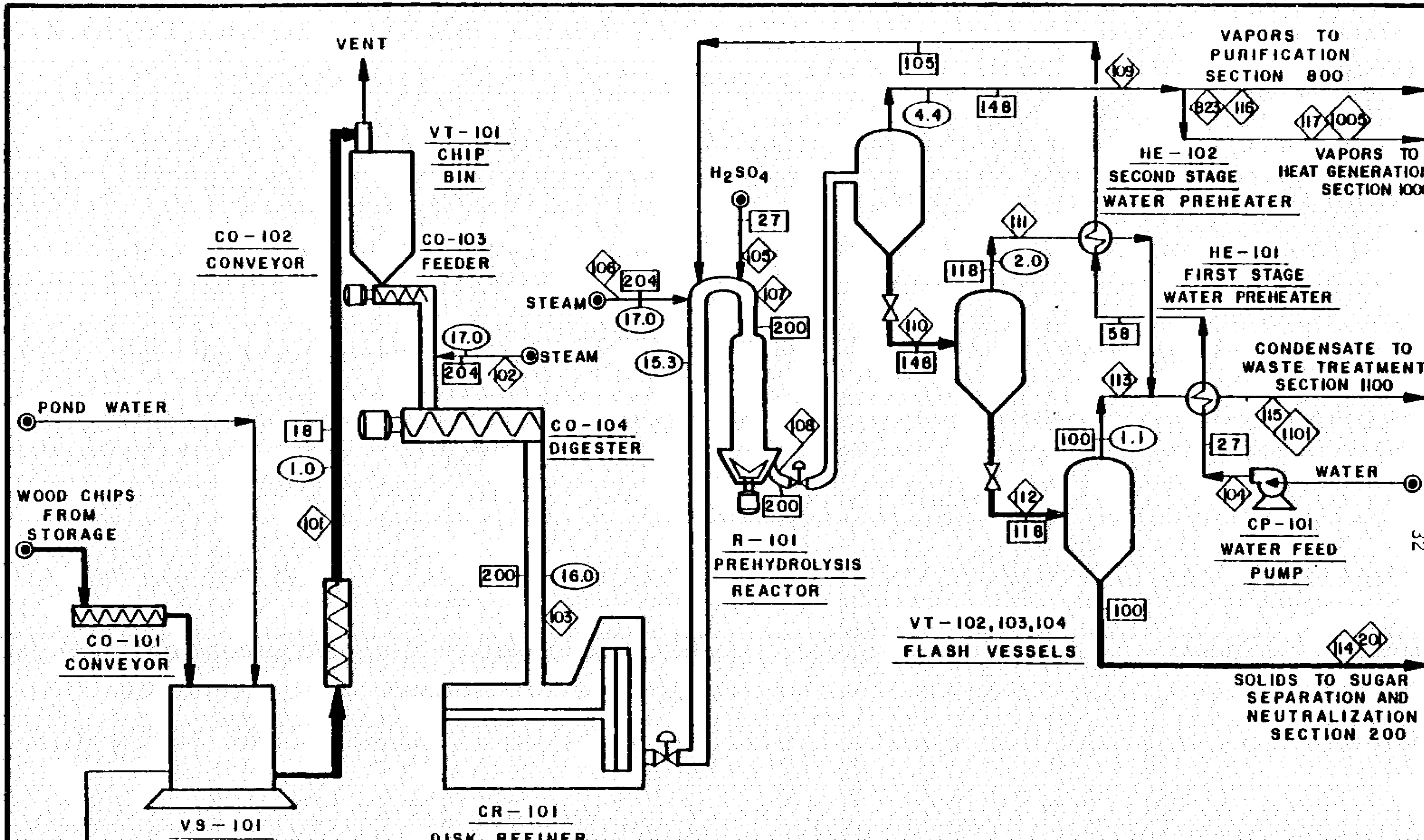

FIGURE II-A-1

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ETHANOL FROM CELLULOSE
SECTION 100
PRETREATMENT/PREHYDROLYSIS

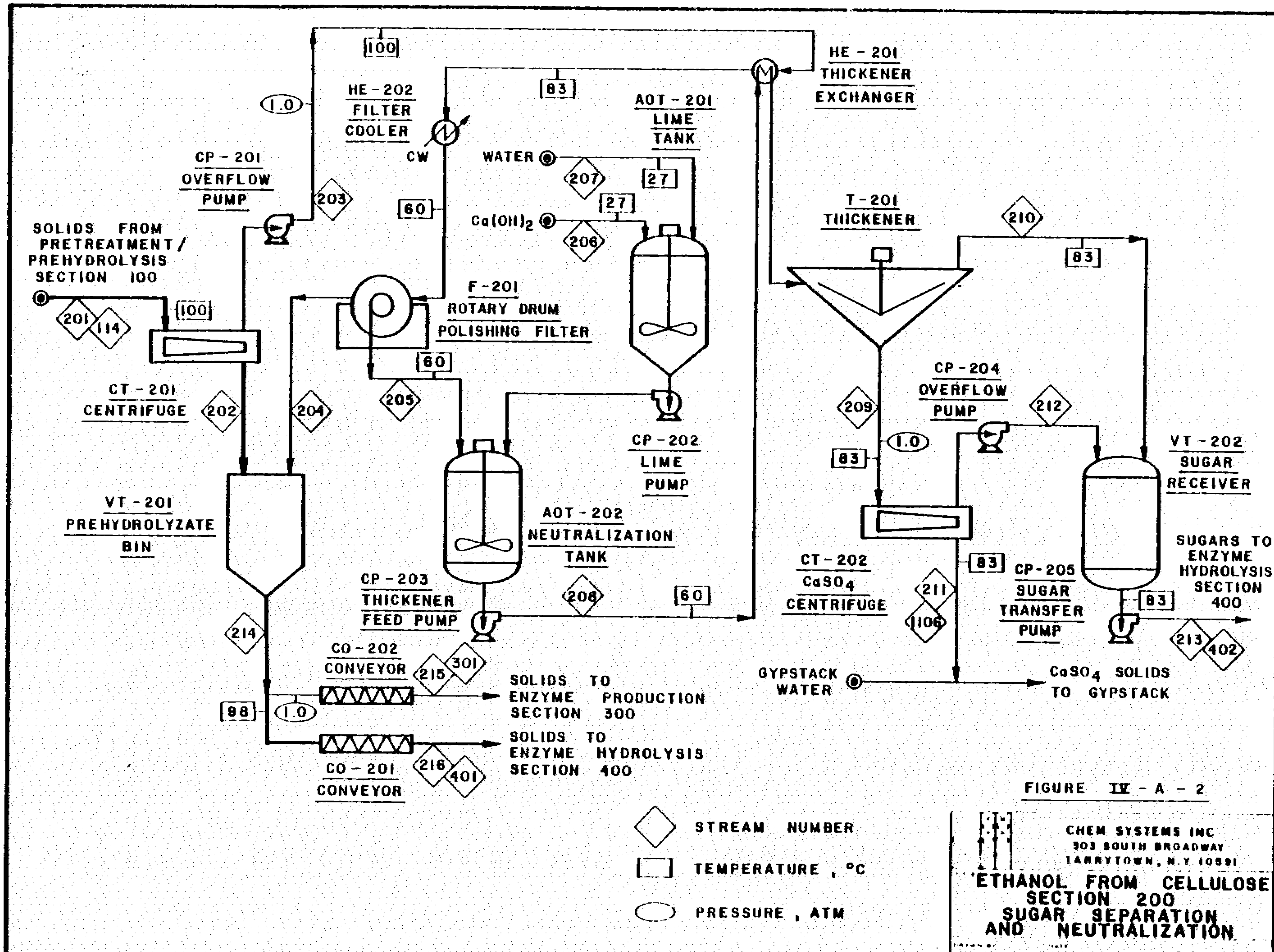
NO. 6042

The prehydrolysis reactor, R-101, is designed for a 12-second residence time at 200°C and 15.3 atm. The wood particles are mixed with water (Stream 104) and steam (Stream 106) to achieve a 15 percent solids concentration and then sulfuric acid is added (Stream 105) to a concentration of 0.75 weight percent based on the water content. The water feed (Stream 104) is preheated by exchange with flash vapors in first stage water preheater HE-101 and in second stage water preheater H-102 to reach a temperature of 105°C prior to injection into the prehydrolysis reactor. All remaining heat is supplied by saturated 17 atm steam in Stream 106. Within the prehydrolysis reactor most of the hemicellulose fraction and the amorphous cellulose are converted to their respective sugars and some degradation products such as furfural. The reactor product (Stream 108) is flashed in three stages to remove some water/furfural vapor and simultaneously recover heat from the prehydrolysis reactor.

The first flash vessel, VT-102, operates at 4.4 atm and 148°C. Liquid/solid discharge (Stream 110) proceeds to the second flash vessel, VT-103, operating at 2.0 atm. Discharge from this flash (Stream 112) is further let down to 1.1 atm in VT-104. The effluent from the third flash (Stream 114), containing approximately 14 weight percent solids, passes to the sugar separation centrifuge in Section 200. Flash vapors (Stream 109) from VT-102 are sent to the multi-effect evaporator steam chest in Section 1000 (Stream 117) and to purification in Section 800 (Stream 116) as beer still preheat. Stream 111 from VT-103 is condensed against prehydrolysis feed water in second stage water preheater HE-102. The condensate is let down and combined with Stream 113 from the third stage flash. This stream is condensed against prehydrolysis feed water in first stage water preheater HE-101. The combined furfural/water condensate (Stream 115) proceeds to waste treatment in Section 1100.

2. Sugar Separation and Neutralization (Section 200) (Figure IV-A-2)

The sugar separation and neutralization section can be subdivided into three subsections:


HE - 201
THICKENER
EXCHANGER
HE - 202
FILTER
COOLER
CW
AOT - 201
LIME
TANK
WATER
Ca(OH)$_2$
CP - 201
OVERFLOW
PUMP
SOLIDS FROM
PRETREATMENT /
PREHYDROLYSIS
SECTION 100
F - 201
ROTARY DRUM
POLISHING FILTER
T - 201
THICKENER
CT - 201
CENTRIFUGE
CP - 202
LIME
PUMP
CP - 204
OVERFLOW
PUMP
VT - 202
SUGAR
RECEIVER
VT - 201
PREHYDROLYZATE
BIN
AOT - 202
NEUTRALIZATION
TANK
CP - 203
THICKENER
FEED PUMP
CT - 202
CaSO$_4$
CENTRIFUGE
CP - 205
SUGAR
TRANSFER
PUMP
SUGARS TO
ENZYME
HYDROLYSIS
SECTION
400
CO - 202
CONVEYOR
SOLIDS TO
ENZYME PRODUCTION
SECTION 300
CO - 201
CONVEYOR
SOLIDS TO
ENZYME HYDROLYSIS
SECTION 400
GYPSTACK
WATER
CaSO$_4$ SOLIDS
TO GYPSTACK
100
83
1.0
60
207
27
203
206
201
114
210
202
204
205
209
212
208
211
1106
213
402
214
215
301
98
216
401
STREAM NUMBER
TEMPERATURE , °C
PRESSURE , ATM
CHEM SYSTEMS INC
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ETHANOL FROM CELLULOSE
SECTION 200
SUGAR SEPARATION
AND NEUTRALIZATION
NO 6042
FIGURE IV - A - 2

- Sugar separation
- Neutralization
- Precipitated salts separation

Prehydrolyzed wood slurry from Section 100 is fed to centrifuge CT-201 to be separated into solid and liquid streams. The overflow (Stream 203) which contains most of the solubles and approximately 5 percent of the solids is pumped to rotary drum polishing filter F-201 via thickener exchanger HE-201 and filter cooler HE-202 which cool the stream to 60°C. The bottoms from CT-201 (Stream 202) containing 35 weight percent solids, are accumulated in prehydrolyzate bin VT-201. In the polishing filter, approximately 95 percent of the solids in Stream 203 are recovered and sent to VT-201 via Stream 204. The clean sugar solution (Stream 205) proceeds to neutralization. Prehydrolyzate solids (Stream 214) are split with 12 percent conveyed to enzyme production in Section 300 via Stream 215 and the remainder (Stream 216) conveyed to enzyme hydrolysis in Section 400.

In neutralization tank AOT-202, the acid in the sugar solution is neutralized with calcium hydroxide solution from lime tank AOT-201 (Streams 206 and 207). The neutralized sugar solution containing insoluble calcium sulfate (Stream 208) is pumped to a thickener via thickener exchanger HE-201 which reheats the stream to 83°C.

In thickener T-201, the calcium sulfate solids are concentrated to 20 weight percent and fed to a solid bowl centrifuge via Stream 209. The clear thickener overflow (Stream 201) is accumulated in sugar receiver VT-202. Calcium sulfate centrifuge CT-202 produces a high solids content salt (Stream 211) which is dumped into a slurry transfer line and transported to the gypstack. Drainage water from the gypstack is accumulated and recirculated through the slurry transfer line to transport the calcium sulfate solids. Overflow from CT-202 (Stream 212) is pumped to sugar receiver VT-202. The accumulated sugar solution from VT-202 is pumped via Stream 213 to enzyme hydrolysis in Section 400.

3. Enzyme Production (Section 300) (Figure IV-A-3)

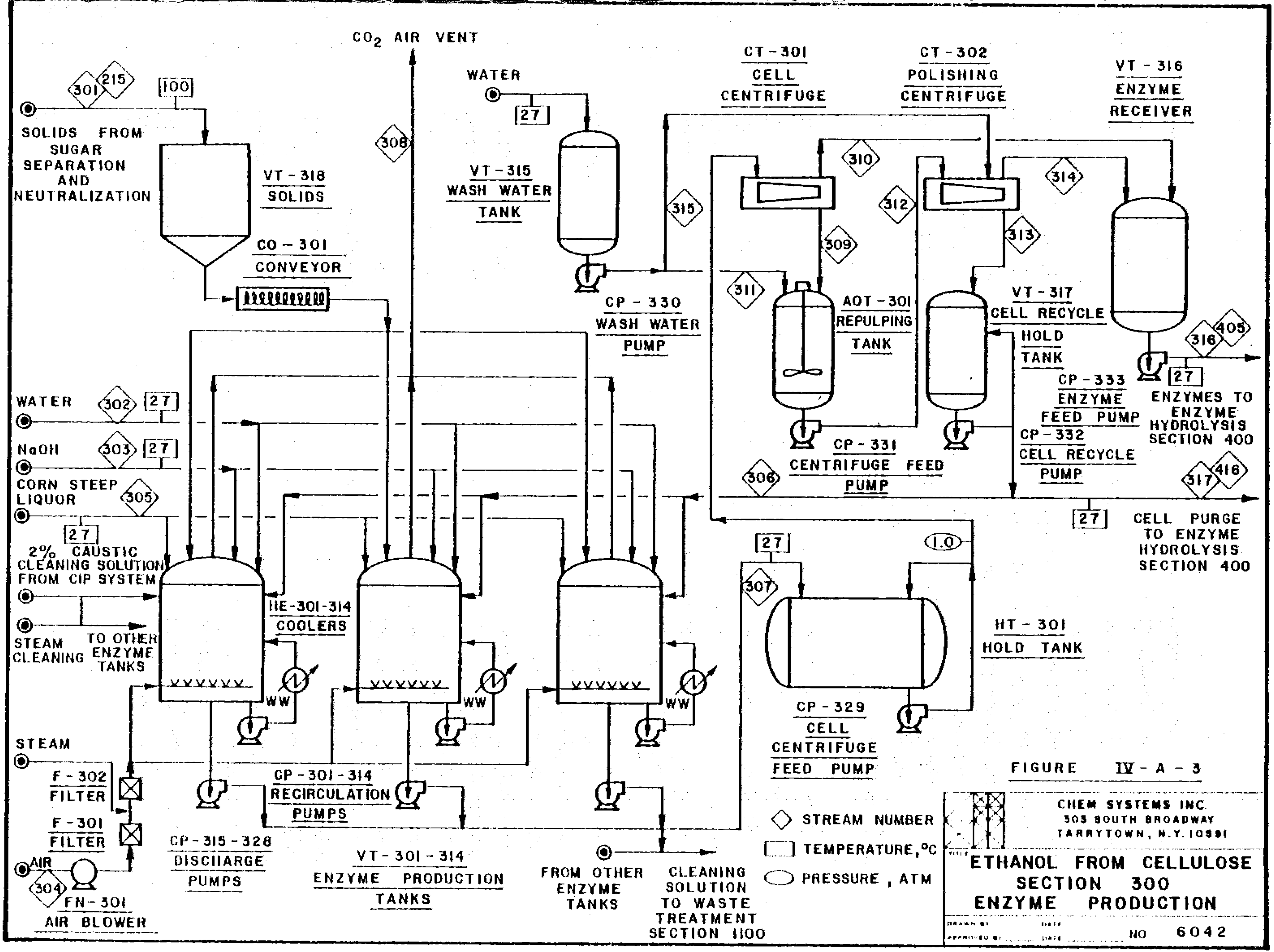
The cellulase enzyme complex, which is used for enzymatic hydrolysis of cellulose, is produced by the microorganism RUT C-30 in a fed batch production system. A portion of the prehydrolyzed wood feedstock is used as the carbon source for enzyme growth. The fed batch system differs from a conventional batch system in that fresh solid feed is added to the enzyme production tanks periodically over the course of the total batch. In this way the actual total solids concentration being processed is much higher than that obtained for a conventional batch system, since fresh solids are periodically added as soon as the previously added solids are consumed.

The enzyme production section can be subdivided into two subsections:

- Enzyme fermentation
- Enzyme separation and cell recycle

The fed batch production system is designed such that the solids feed stream is essentially continuous. Thirteen separate enzyme fermentation tanks are used, with one additional tank being out of operation as a reserve. The total batch cycle time for each tank is 13 days or 312 hours, including emptying and sterilization for the next batch. One batch is started each day with filling taking four hours. Fresh solids are added to each tank approximately every 48 hours or six times during the entire batch cycle. One batch is emptied each day after 296 hours on stream. Emptying takes 4 hours and then the remaining 12 hours before starting the next batch are used for sterilization and cleaning. Dilution water, nutrients, and cell recycle are added only once for each cycle during the initial filling of the tank. Fresh solids and caustic for neutralization are added six times during each cycle.

Solids from sugar separation and neutralization (Section 200) which serve as the carbon source for enzyme growth, are continuously fed to solids bin, VT-318 (Steam 301) prior to being fed to the individual enzyme production tanks. The solids are then fed to the enzyme production tanks,


CO2 AIR VENT
SOLIDS FROM SUGAR SEPARATION AND NEUTRALIZATION
VT - 318 SOLIDS
CO - 301 CONVEYOR
WATER
VT - 315 WASH WATER TANK
CP - 330 WASH WATER PUMP
CT - 301 CELL CENTRIFUGE
CT - 302 POLISHING CENTRIFUGE
VT - 316 ENZYME RECEIVER
AOT - 301 REPULPING TANK
VT - 317 CELL RECYCLE HOLD TANK
CP - 333 ENZYME FEED PUMP
CP - 332 CELL RECYCLE PUMP
CP - 331 CENTRIFUGE FEED PUMP
ENZYMES TO ENZYME HYDROLYSIS SECTION 400
CELL PURGE TO ENZYME HYDROLYSIS SECTION 400
WATER
NaOH
CORN STEEP LIQUOR
2% CAUSTIC CLEANING SOLUTION FROM CIP SYSTEM
STEAM CLEANING
TO OTHER ENZYME TANKS
HE-301-314 COOLERS
WW
HT - 301 HOLD TANK
CP - 329 CELL CENTRIFUGE FEED PUMP
STEAM
F - 302 FILTER
F - 301 FILTER
AIR
FN - 301 AIR BLOWER
CP - 315 - 328 DISCHARGE PUMPS
CP - 301 - 314 RECIRCULATION PUMPS
VT - 301 - 314 ENZYME PRODUCTION TANKS
FROM OTHER ENZYME TANKS
CLEANING SOLUTION TO WASTE TREATMENT SECTION 1100
STREAM NUMBER
TEMPERATURE, °C
PRESSURE, ATM
FIGURE IV - A - 3
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ETHANOL FROM CELLULOSE
SECTION 300
ENZYME PRODUCTION
NO 6042

VT-301-313, in the manner previously described, along with nutrient corn steep liquor (Stream 305), recycle cells (Stream 306), caustic (Stream 303), and dilution water (Stream 302). The aerobic fermentation oxygen requirement is met by sterilized air (Stream 304) being continuously sparged through the enzyme production tanks via air blower FN-301 and filters F-301 and F-302 which also provides agitation in the tanks. Enzyme production pH is controlled at 5.0 by the addition of the caustic, and temperature is controlled at 28^{o}C by an external cooling line through heat exchangers HE-301-314. Excess air, nitrogen and the CO_2 formed during enzyme production are vented from the tanks to the atmosphere (Stream 308).

Initial cellulose concentration in the fermenters is about 1.7 weight percent, and total solids concentration, including recycle solids, is about 9.4 percent. Cell density is initially maintained at 7 grams per liter, and inital corn steep liquor concentration is 3 weight percent. Final enzyme concentration is 15 IU/ml which translates to an enzyme productivity of about 52 IU/liter/hour.

The enzyme fermenter effluent (Stream 307) is discharged from the tanks via pumps CP-315-328 to hold tank, HT-301. The enzyme solution is then fed to the solid bowl cell centrifuge, CT-301, where the mycellium and other solids are removed from the enzyme solution. The centrifuge bottoms (Stream 309), which is 15 weight percent solids, is repulped to 10 percent solids by addition of water (Stream 311) in repulping tank, AOT-301. The repulped solids (Stream 312) are sent to a washing polishing centrifuge, CT-302, where solids are once again removed as the bottoms (Stream 313), and sent to the cell recycle hold tank, VT-317. The overflow from the polishing centrifuge (Stream 314) is combined with the overflow from the cell centrifuge (Stream 310) in the enzyme receiver, VT 316, and sent to enzyme hydrolysis, Section 400 (Stream 316). The cells are discharged from the cell recycle hold tank via pump CP-332, part of which is recycled to the enzyme production tanks (Stream 306). The remaining cells are purged (Stream 317) to the dewatering press in enzyme hydrolysis, Section 400. The cell recycle rate is approximately 55 percent of the total cell stream.

4. Enzyme Hydrolysis (Section 400) (Figure IV-A-4)

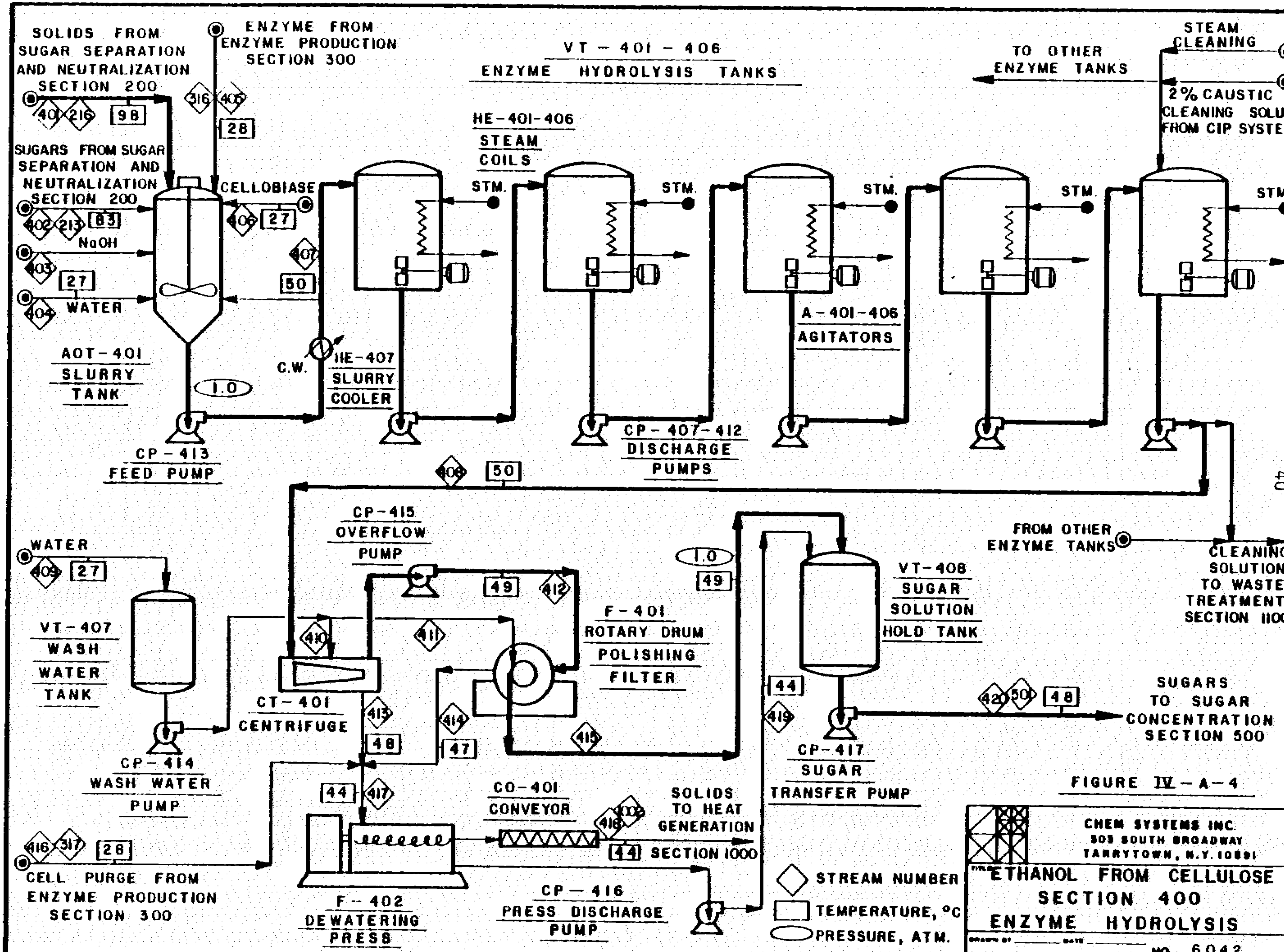
The enzyme hydrolysis section can be subdivided into two subsections:

- Enzyme hydrolysis
- Sugar separation

The solid stream (Stream 401) and sugar stream (Stream 402) from Section 200, sugar separation and neutralization, are fed to slurry tank, AOT-401. Slurry pH is controlled at 4.8 by addition of sodium hydroxide/calcium hydroxide (Stream 403) such that the final slurry is slightly below the saturation point of calcium sulfate. Cellulase enzymes from Section 300, enzyme production, (Stream 405) and make up NOVO cellobiase enzyme (Stream 406) are also added to the slurry, along with sufficient dilution water (Stream 404) to bring the total solids concentration to 10 weight percent. Temperature is controlled at 50°C via an external cooling line on the slurry tank through HE-407.

Enzyme hydrolysis takes place in a continuous cascade system in which partial hydrolysis occurs in each of a series of five tanks, until hydrolysis is complete in the last tank. The pH controlled enzyme hydrolysis feed (Stream 407) is pumped from the slurry tank to the enzyme hdyrolysis tanks, VT-401-406. Enzyme hydrolysis takes place at 50°C with a cellulase loading of 15 IU per gram of solids and supplemented by a cellobiase loading of 0.9 IU per gram of solids. Total hydrolysis residence time in the five tanks is 24 hours. Agitation is provided in the tanks by internal side-mounted agitators and temperature is controlled by internal heating/cooling coils. The enzyme hydrolysis conditions give a 90 percent of theoretical glucose yield with 10 percent cellulose remaining unconverted. All remaining xylan present in the hydrolysis feed is assumed to be converted to xylose.

The enzyme hydrolysis effluent (Stream 408), which contains about 6.9 percent glucose is sent to a series of centrifuging and filtering steps. The first separation takes place in a washing centrifuge, CT-401, where most of the lignin and other remaining solids are removed in the


SOLIDS FROM SUGAR SEPARATION AND NEUTRALIZATION SECTION 200
ENZYME FROM ENZYME PRODUCTION SECTION 300
VT – 401 – 406 ENZYME HYDROLYSIS TANKS
TO OTHER ENZYME TANKS
STEAM CLEANING
2% CAUSTIC CLEANING SOLUT. FROM CIP SYSTEM
HE-401-406 STEAM COILS
SUGARS FROM SUGAR SEPARATION AND NEUTRALIZATION SECTION 200
CELLOBIASE
NaOH
WATER
STM.
AOT-401 SLURRY TANK
C.W.
HE-407 SLURRY COOLER
A-401-406 AGITATORS
CP-413 FEED PUMP
CP-407-412 DISCHARGE PUMPS
CP-415 OVERFLOW PUMP
FROM OTHER ENZYME TANKS
CLEANING SOLUTION TO WASTE TREATMENT SECTION 1100
VT-407 WASH WATER TANK
F-401 ROTARY DRUM POLISHING FILTER
VT-408 SUGAR SOLUTION HOLD TANK
CT-401 CENTRIFUGE
CP-414 WASH WATER PUMP
SUGARS TO SUGAR CONCENTRATION SECTION 500
CP-417 SUGAR TRANSFER PUMP
CO-401 CONVEYOR
SOLIDS TO HEAT GENERATION SECTION 1000
CELL PURGE FROM ENZYME PRODUCTION SECTION 300
F-402 DEWATERING PRESS
CP-416 PRESS DISCHARGE PUMP
STREAM NUMBER
TEMPERATURE, °C
PRESSURE, ATM.
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ETHANOL FROM CELLULOSE
SECTION 400
ENZYME HYDROLYSIS
NO. 6042

FIGURE IV – A – 4

centrifuge bottoms (Stream 413). The centrifuge overflow (Stream 412) is sent to a rotary drum polishing filter, F-401, where most of the remaining solids are removed. The bottoms from the centrifuge and filter, which contain about 35 percent solids, are combined with the cell purge (Stream 416) from enzyme production, Section 300, and sent to dewatering press F-402 (Stream 417). The dewatering press squeezes out additional liquid to bring the solid concentration to about 55 percent. This solid stream (Stream 418) is sent to heat generation, Section 1000. The filtrate from the polishing filter (Stream 415) and the liquid from the dewatering press (Stream 419) are sent to the sugar solution hold tank, VT-408, prior to concentration. The feed to the sugar concentrator, Section 500 (Stream 420) contains about 6.8 percent glucose.

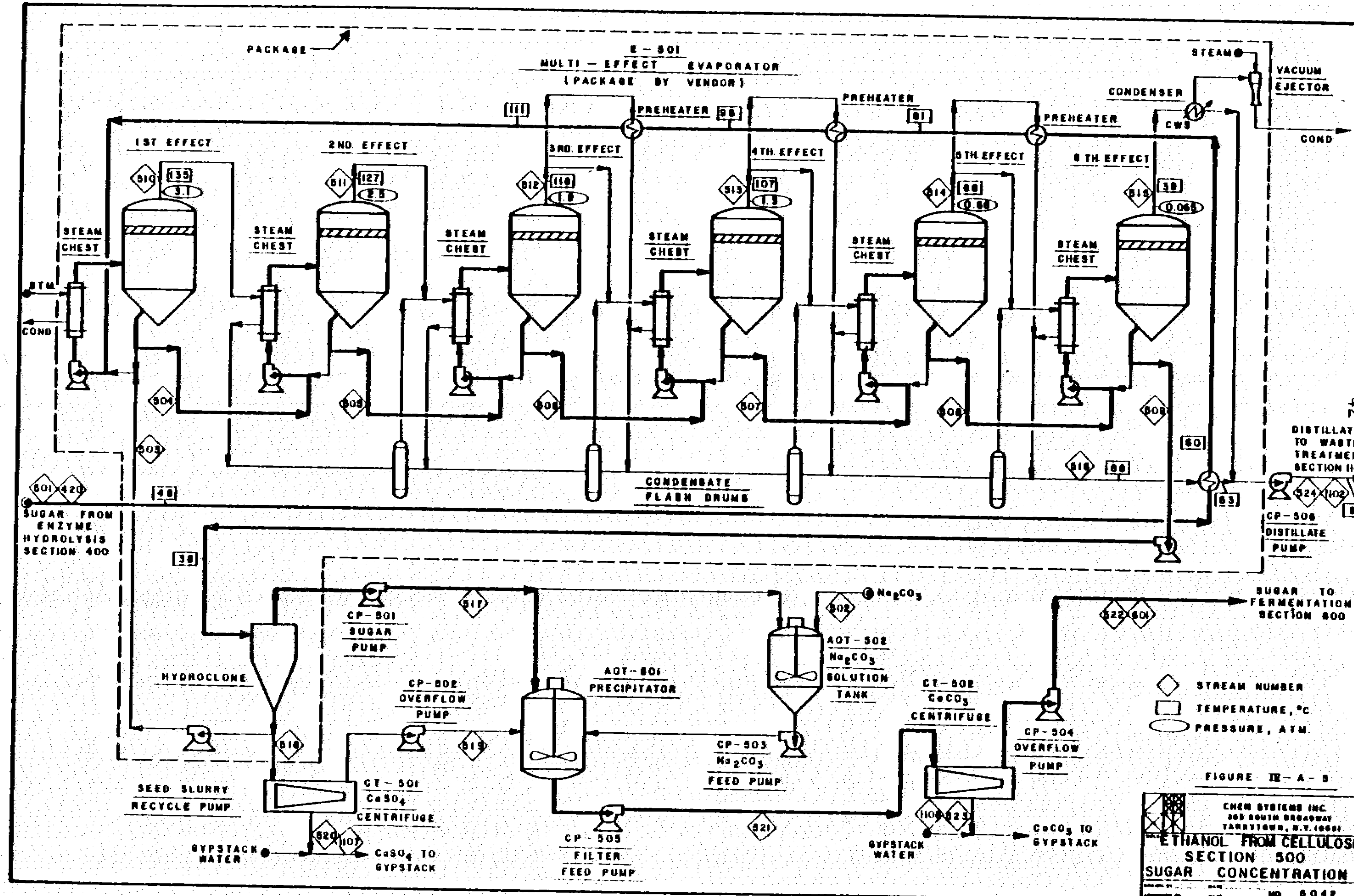
5. Sugar Concentration (Section 500) (Figure IV-A-5)

The sugar concentration can be subdivided into two subsections:

- Sugar concentration
- Salt separation

The 6.8 weight percent sugar solution from enzyme hydrolysis is concentrated to 15 weight percent in multi-effect evaporator E-501. This evaporator is a six-effect, external forward forced fed system, constructed of 304 SS based on vendor recommendations for this service. The first effect operates at 3.1 atm and the final effect under vacuum at 0.065 atm.

Several heat recovery steps are employed to minimize the overall steam duty. These include the flash of the steam chest condensate after each effect such that total distillate from effects two through five has been reduced to 88°C (0.66 atm). Also the cold feed (50°C) is preheated in a series of exchangers to 111°C. The flowsheet depicts a complete vendor package for the evaporator including heat recovery equipment. In the operation of the concentrator system the feed (Stream 501) is first heat exchanged against the 88°C distillate (Stream 516) then further heated to 111°C using an overhead vapor bleed from the fifth, fourth,



PACKAGE
E - 501
MULTI - EFFECT EVAPORATOR
(PACKAGE BY VENDOR)
1ST EFFECT
2ND. EFFECT
3RD. EFFECT
4TH. EFFECT
5TH EFFECT
6TH EFFECT
PREHEATER
PREHEATER
PREHEATER
CONDENSER
STEAM
VACUUM EJECTOR
CWS
COND
STEAM CHEST
STM
COND
CONDENSATE FLASH DRUMS
SUGAR FROM ENZYME HYDROLYSIS SECTION 400
DISTILLATE TO WASTE TREATMENT SECTION 1100
CP-506 DISTILLATE PUMP
HYDROCLONE
CP-501 SUGAR PUMP
AGT-501 PRECIPITATOR
AGT-502 Na_2CO_3 SOLUTION TANK
Na_2CO_3
SUGAR TO FERMENTATION SECTION 600
CP-502 OVERFLOW PUMP
CP-503 Na_2CO_3 FEED PUMP
CT-502 $CaCO_3$ CENTRIFUGE
CP-504 OVERFLOW PUMP
SEED SLURRY RECYCLE PUMP
CT-501 $CaSO_4$ CENTRIFUGE
GYPSTACK WATER
$CaSO_4$ TO GYPSTACK
CP-505 FILTER FEED PUMP
GYPSTACK WATER
$CaCO_3$ TO GYPSTACK
STREAM NUMBER
TEMPERATURE, °C
PRESSURE, ATM.
FIGURE IV-A-5
CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591
ETHANOL FROM CELLULOSE
SECTION 500
SUGAR CONCENTRATION
NO 6042

and third effects, respectively. This scheme serves to reduce the amount of unrecoverable heat from the steams (i.e., the 6th effect distillate, Stream 515, is vapor at 38°C).

The hot feed then enters the first effect. In each of the next effects the exchanger is driven by the overhead of the previous effect. Also, the condensed distillate from each steam chest is flashed and fed to the next heater for the purpose of heat recovery.

Since the feed stream is saturated with calcium sulfate, this salt will precipitate as the liquor is concentrated. To minimize fouling of the exchangers, a crystal seeding process is employed. This is accomplished by feeding the liquor out of the sixth effect to a hydroclone where a clear solution is taken overhead (Stream 517) and a 40 weight percent calcium sulfate stream is taken off the bottom. A portion of the bottoms is recycled back to the first effect such that the final effect on the evaporator contains an 8 weight percent slurry of calcium sulfate salt. This provides seeds on which the precipitating solids build rather than plating on the equipment walls. As this system does not totally prevent fouling, the heat exchangers must be cleaned periodically.

A slipstream (Stream 518) from the hydroclone bottoms, containing the net amount of calcium sulfate in the feed (Stream 501), is sent to the calcium sulfate centrifuge, CT-501. The solids stream containing 77 weight percent organic and inorganic solids (Stream 520) is sent to the gypstack.

The liquid from CT-501 (Stream 519) along with the clear liquid from the hydroclone (Stream 517) is pumped to the precipitator, AOT-501, where sodium carbonate is added such that the calcium sulfate is converted to calcium carbonate. The reason for calcium carbonate conversion is that this salt is much less soluble than calcium sulfate. Since both have reverse solubilities, there will be less calcium salts that can plate out in the downstream beerstill as the stream is heated. The insoluble salts are separated in the calcium carbonate centrifuge, CT-502. The solids bottoms containing 90 weight percent calcium carbonate (Stream 523) are sent to the gypstack while the liquor is sent to fermentation in Section 600 (Stream 522).

6. Fermentation (Section 600) (Figure IV-A-6)

The 15 percent sugar solution from sugar concentration in Section 500 (Stream 601) is fed to concentrated sugar receiver VT-608. Sterile aeration is provided to this vessel through air blower FN-602 with filters F-601 and F-602 preventing external contamination. The sugar solution (Stream 603) is pumped through sugar cooler HE-607 into the fermenters. Recycle yeast is also added to the fermenters via stream 605.

Fermentation takes place in a continuous cascade system consisting of six tanks, VT-601-606, with five tanks active at any one time and the additional tank down for sterilization. Partial fermentation occurs in each tank and is completed in the last tank. The total fermentation time is 18 hours. The yeast, Saccharomyces cerevisiae, is fed to fermentation at a concentration of 385 x 10^6 cells per millimeter or approximately 0.07 pounds per gallon. The pH in fermentation is adjusted to 4.0 primarily by addition of sulfuric acid to the yeast recycle stream. Well water cooling is provided to each fermenter through external coolers, HE-601-606 to maintain a temperature of 30°C.

In fermentaion, all of the glucose is converted with 93 percent selectivity to ethanol and carbon dioxide; 4.9 percent selectivity to glycerol, acetaldehyde, and carbon dioxide; 0.1 percent selectivity to fusel oils; and 2 percent selectivity to yeast growth and carbon dioxide production. Thus, the ethanol productivity is 0.475 pounds per pound of glucose and 0.476 pounds of carbon dioxide are produced per pound of glucose. Yeast growth is approximately 18 percent per pass. Thus, the entire yeast inventory can be changed out in less than a week, even without fresh yeast addition. Under these conditions, no yeast makeup is necessary.

Carbon dioxide evolved during fermentation (Stream 607) passes through scrubber TW-601 where entrained ethanol is absorbed in water from stream 608, pumped via stream 602 to concentrated sugar receiver VT-608 and then returned to the fermenters. The scrubbed carbon dioxide (Stream 609) is sent to carbon dioxide recovery in Section 700.

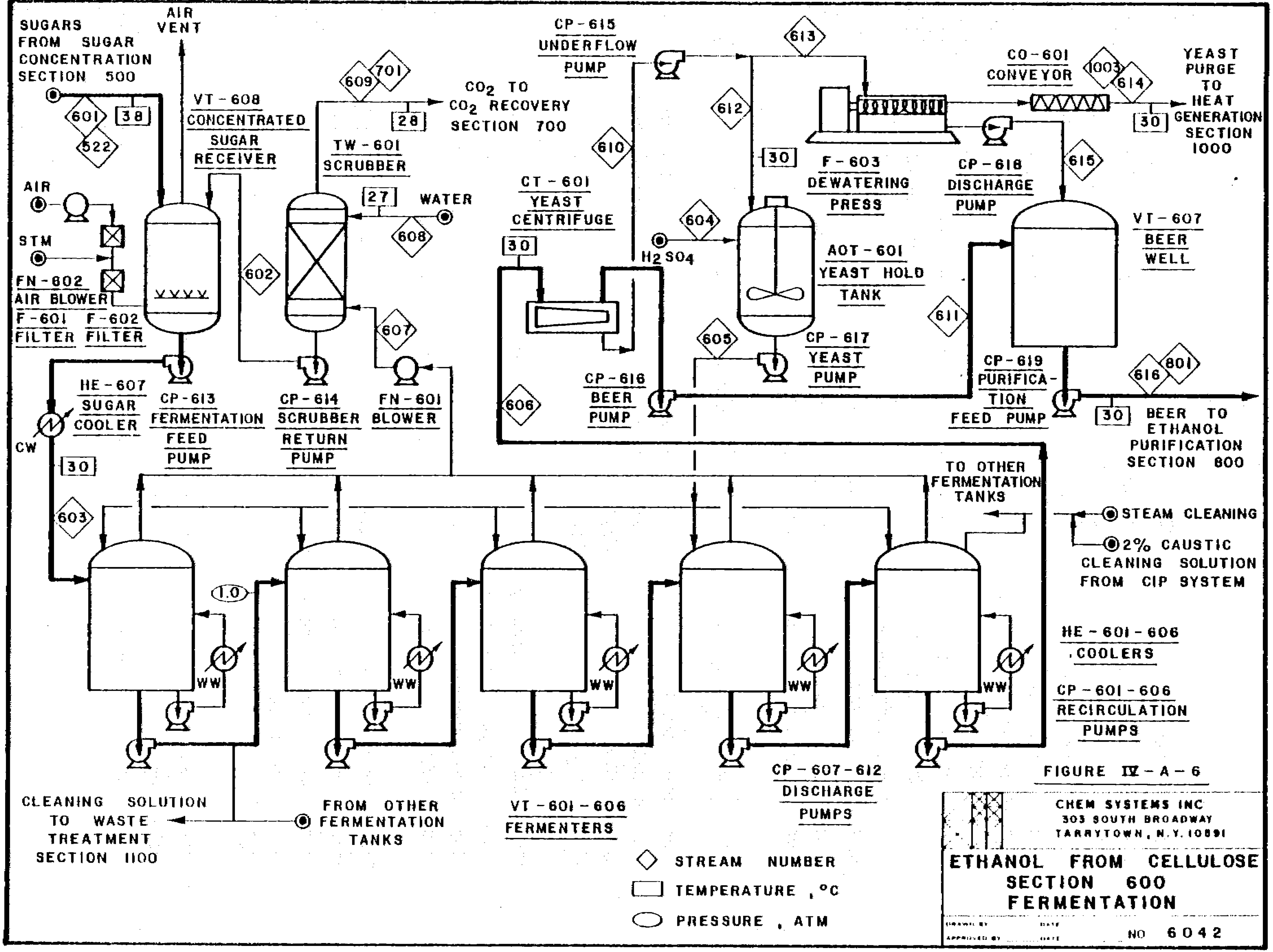

SUGARS FROM SUGAR CONCENTRATION SECTION 500
AIR VENT
VT - 608 CONCENTRATED SUGAR RECEIVER
CO2 TO CO2 RECOVERY SECTION 700
TW - 601 SCRUBBER
CP - 615 UNDERFLOW PUMP
CO - 601 CONVEYOR
YEAST PURGE TO HEAT GENERATION SECTION 1000
F - 603 DEWATERING PRESS
CP - 618 DISCHARGE PUMP
AIR
STM
FN - 602 AIR BLOWER
F - 601 FILTER
F - 602 FILTER
WATER
CT - 601 YEAST CENTRIFUGE
H2 SO4
AOT - 601 YEAST HOLD TANK
VT - 607 BEER WELL
HE - 607 SUGAR COOLER
CW
CP - 613 FERMENTATION FEED PUMP
CP - 614 SCRUBBER RETURN PUMP
FN - 601 BLOWER
CP - 616 BEER PUMP
CP - 617 YEAST PUMP
CP - 619 PURIFICATION FEED PUMP
BEER TO ETHANOL PURIFICATION SECTION 800
TO OTHER FERMENTATION TANKS
STEAM CLEANING
2% CAUSTIC CLEANING SOLUTION FROM CIP SYSTEM
HE - 601 - 606 COOLERS
CP - 601 - 606 RECIRCULATION PUMPS
WW
CP - 607 - 612 DISCHARGE PUMPS
CLEANING SOLUTION TO WASTE TREATMENT SECTION 1100
FROM OTHER FERMENTATION TANKS
VT - 601 - 606 FERMENTERS
601 522 38 28 609 701 27 608 602 607 603 30 1.0 613 612 610 604 605 606 611 615 614 1003 616 801
STREAM NUMBER
TEMPERATURE , °C
PRESSURE , ATM
CHEM SYSTEMS INC
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591
ETHANOL FROM CELLULOSE
SECTION 600
FERMENTATION
NO 6042

FIGURE IV - A - 6

The beer exiting the fermentation tanks (Stream 606) has an ethanol concentration of 7.0 weight percent. This stream is sent to yeast centrifuge CT-601 where a 15 weight percent yeast (Stream 610) is separated from the ethanol containing beer (Stream 611). The beer is pumped to beer well VT-607. Approximately 90 percent of the yeast (Stream 611) is sent to yeast hold tank AOT-601 where sulfuric acid is added via stream 604 to bring the pH to 3. The yeast is held at this condition for four hours to kill off bacteria and then recycled to the fermenters via stream 605. A yeast purge (Stream 613) is sent to dewatering press F-603 where the solids concentration is raised to 55 weight percent. These solids (Stream 614) are conveyed to heat generation in Section 1000. The dewatered liquids (Stream 615) are sent to the beer well. From VT-607 the beer (Stream 616) is pumped to purification in Section 800.

7. Carbon Dioxide Recovery (Section 700) (Figure IV-A-7)

When sugars are fermented to ethanol, large quantities of carbon dioxide are produced as a by-product of cell respiration. As the fermenters are usually closed vessels, it is possible to collect the off-gas and recover liquefied carbon dioxide for sale.

The carbon dioxide generated in the fermentation vessels (Stream 701) passes through a foam trap, boosted in pressure with a rotary positive displacement compressor and discharged to a surge vessel. The gas is then compressed to 21.4 atm in a nonlubricated, reciprocating compressor.

The compressed gas is deodorized in a twin-tower, activated-carbon absorption system to remove remaining impurities. The carbon beds are periodically regenerated using live steam. The purified gas is then chilled and dried in a conventional alumina bed system to a dew point of -60°C.

The dry gas passes to a low-temperature stripper-condenser system, where the carbon dioxide is liquefied and separated from the non-condensible gases, mainly nitrogen, which are vented to the atmosphere. The pure

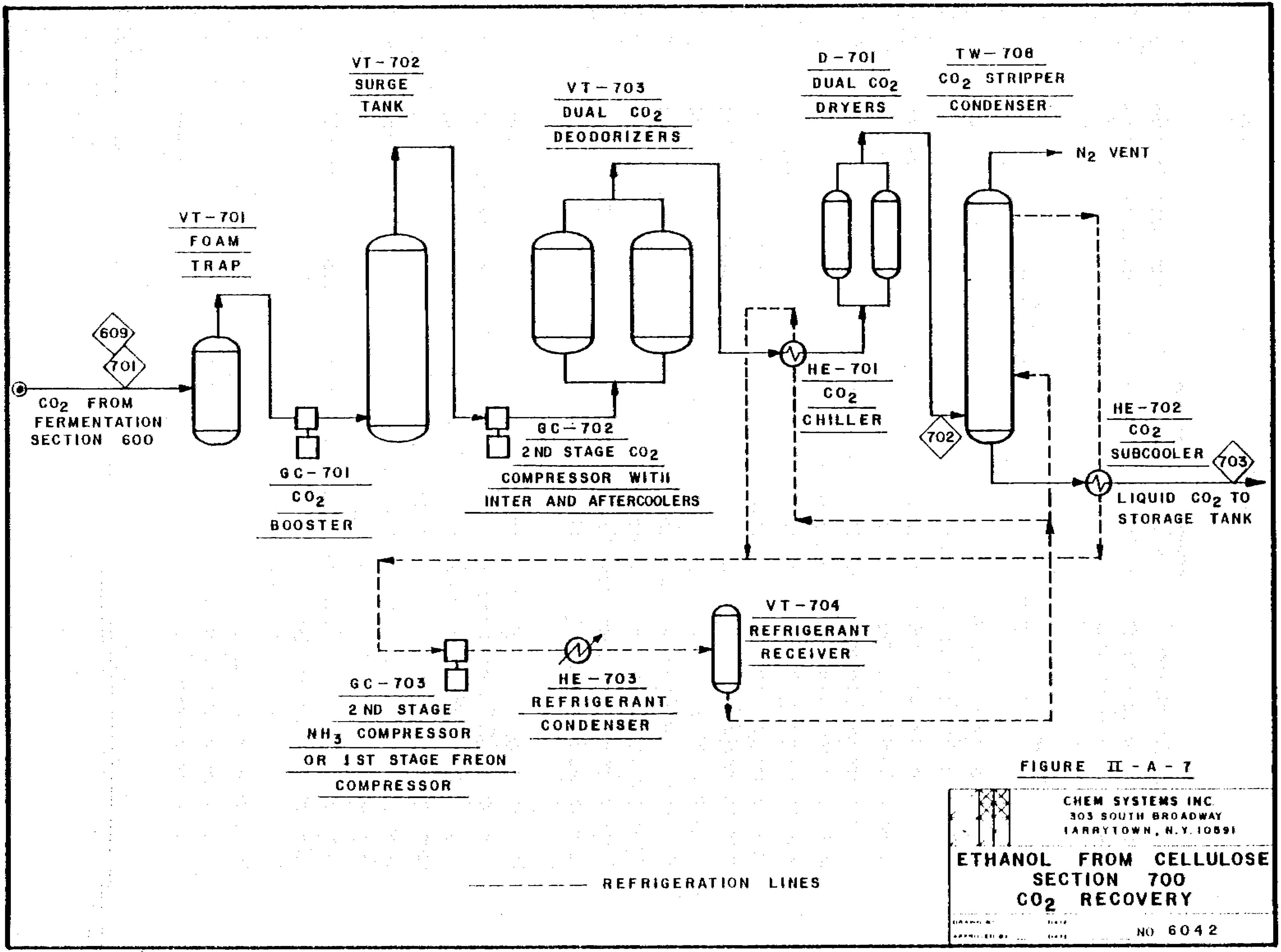

VT - 701
FOAM
TRAP
VT - 702
SURGE
TANK
VT - 703
DUAL CO_2
DEODORIZERS
D - 701
DUAL CO_2
DRYERS
TW - 708
CO_2 STRIPPER
CONDENSER
N_2 VENT
609
701
CO_2 FROM
FERMENTATION
SECTION 600
GC - 701
CO_2
BOOSTER
GC - 702
2 ND STAGE CO_2
COMPRESSOR WITH
INTER AND AFTERCOOLERS
HE - 701
CO_2
CHILLER
702
HE - 702
CO_2
SUBCOOLER
703
LIQUID CO_2 TO
STORAGE TANK
VT - 704
REFRIGERANT
RECEIVER
GC - 703
2 ND STAGE
NH_3 COMPRESSOR
OR 1 ST STAGE FREON
COMPRESSOR
HE - 703
REFRIGERANT
CONDENSER
REFRIGERATION LINES
CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591
ETHANOL FROM CELLULOSE
SECTION 700
CO_2 RECOVERY
NO 6042

FIGURE II - A - 7

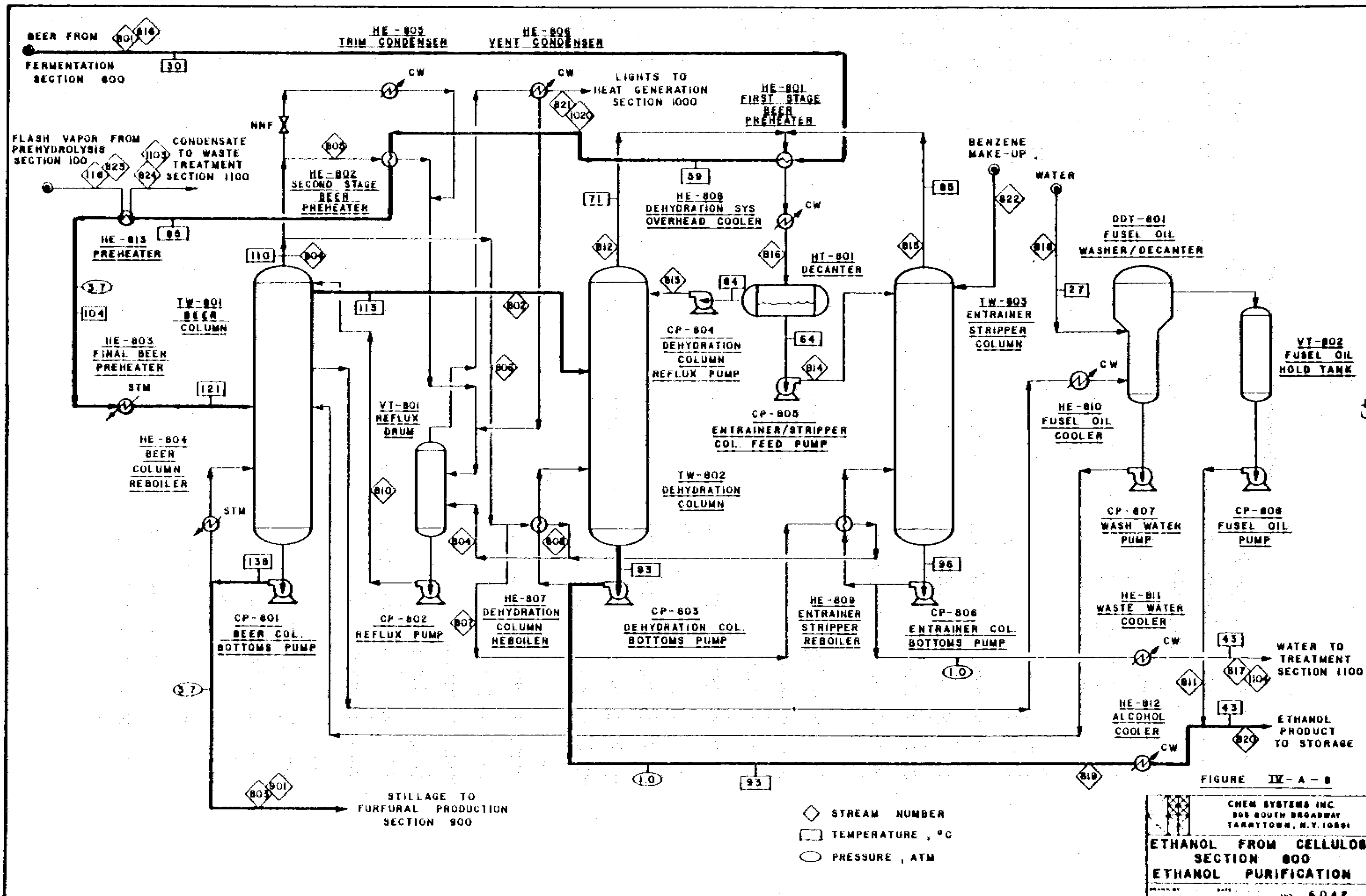
liquid carbon dioxide from the base of the stripper-condenser is then subcooled and sent to storage (Stream 703), where it is maintained under a pressure of about 21.4 atm.

8. Ethanol Purification (Section 800) (Figure IV-A-8)

The dilute beer from fermentation (Stream 801) undergoes a series of preheating steps that heat the feed from 30°C to 121°C. The first preheating step utilizes the combined overhead vapors from the dehydration column, TW-802, and from the entrainer stripper column, TW-803, in the first-stage beer preheater, HE-801. Approximately 28 percent of the total preheating is accomplished here, bringing the beer up to 59°C. The warmed beer then enters the second-stage beer preheater, HE-802, which uses the beer column overheads to supply another 30 percent of preheat, raising the beer temperature to 85°C. The next 22 percent of the preheat is supplied in preheater HE-813 by exchange against flash vapors from Section 100 (Stream 823) which raises the temperature to 104°C. From here the beer is preheated with steam in final beer preheater HE-803. This supplies the final 20 percent of beer preheat, raising the temperature to 121°C. All four stages of preheating use condensing vapors.

The hot, saturated dilute beer then enters the beer column, TW-801, which operates at 3.7 atm at the bottom. The column is composed of 60 sieve trays. The nonvolatile solubles and any suspended solids work their way down through the column and are removed from the bottom of the tower in a dilute aqueous stream. The temperature of the bottoms is 138°C. The hot stillage (Stream 803) proceeds to the reactor in furfural production (Section 900). Heat is supplied to the beer column by condensing steam in beer column reboiler HE-804.

The alcohol-rich vapors in the overhead of the beer column (Stream 804) exit as the binary azeotrope (94 percent ethanol and 6 percent water) at a temperature of about 110°C and a pressure of 3.4 atm. These vapors are



BEER FROM FERMENTATION SECTION 600
801
814
30
HE-805 TRIM CONDENSER
HE-806 VENT CONDENSER
CW
LIGHTS TO HEAT GENERATION SECTION 1000
821
1020
HE-801 FIRST STAGE BEER PREHEATER
NNF
FLASH VAPOR FROM PREHYDROLYSIS SECTION 100
CONDENSATE TO WASTE TREATMENT SECTION 1100
1103
116
823
824
805
HE-802 SECOND STAGE BEER PREHEATER
59
HE-808 DEHYDRATION SYS OVERHEAD COOLER
71
85
BENZENE MAKE-UP
822
WATER
HE-813 PREHEATER
95
110
804
812
816
815
818
HT-801 DECANTER
DDT-801 FUSEL OIL WASHER/DECANTER
3.7
104
TW-801 BEER COLUMN
113
802
813
84
64
CP-804 DEHYDRATION COLUMN REFLUX PUMP
TW-803 ENTRAINER STRIPPER COLUMN
27
VT-802 FUSEL OIL HOLD TANK
HE-803 FINAL BEER PREHEATER
806
814
STM
121
CP-805 ENTRAINER/STRIPPER COL. FEED PUMP
HE-810 FUSEL OIL COOLER
HE-804 BEER COLUMN REBOILER
VT-801 REFLUX DRUM
TW-802 DEHYDRATION COLUMN
810
CP-807 WASH WATER PUMP
CP-808 FUSEL OIL PUMP
804
808
138
93
96
CP-801 BEER COL. BOTTOMS PUMP
CP-802 REFLUX PUMP
807
HE-807 DEHYDRATION COLUMN REBOILER
CP-803 DEHYDRATION COL. BOTTOMS PUMP
HE-809 ENTRAINER STRIPPER REBOILER
CP-806 ENTRAINER COL. BOTTOMS PUMP
HE-811 WASTE WATER COOLER
43
WATER TO TREATMENT SECTION 1100
817
1104
811
1.0
HE-812 ALCOHOL COOLER
ETHANOL PRODUCT TO STORAGE
820
819
803
901
STILLAGE TO FURFURAL PRODUCTION SECTION 900
STREAM NUMBER
TEMPERATURE, °C
PRESSURE, ATM
FIGURE IV-A-8
CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591
ETHANOL FROM CELLULOSE
SECTION 800
ETHANOL PURIFICATION
NO 6042

used to preheat the beer in the second-stage beer preheat (Stream 805) and as a source of heat for the reboilers of the dehydration tower (Stream 807) and entrainer stripper column (Stream 808), HE-807 and HE-809, respectively. Of the total overhead vapor generated, 40 percent is used for beer preheating, 57 percent is used in HE-807, and 13 percent is used in HE-809. The total condensate from these operations is collected in reflux drum VT-801 and returned to the top tray of the beer column (Stream 810) as reflux. The light (noncondensable) products of fermentation (Stream 821) such as aldehydes are removed from the system by venting from the reflux drum and sent to heat generation in Section 1000. These vapors enter vent condenser HE-806 to recover any alcohol entrained with them before they exit the distillation system.

The upper trays of the beer column operate under total reflux with the liquid product removed as a side draw about five trays from the top of the tower. This stream, which contains 92.9 percent ethanol and 7.1 percent water, then enters the mid-section of dehydration tower TW-802.

Also removed from the beer column are the fusel oils produced in fermentation. These higher alcohols are more volatile than ethanol in dilute aqueous solutions but less volatile than ethanol in concentrated alcohol solutions. As a result, they tend to concentrate on a tray in the beer column and must be removed. They are removed as a side draw, cooled in fusel oil cooler HE-810 and washed with cold water in the fusel oil washer/decanter, DDT-801. In this operation, the alcohol content of the fusel oils is washed off by countercurrent contact of the cooled fusel oils with a stream of cold water. The heavy aqueous stream, containing the recovered alcohol, is returned to the lower section of the beer column. The light organic (fusel oil) stream passes into fusel oil hold tank VT-802 for later blending with the ethanol product since it is to be used as a motor-grade fuel.

The dehydration system consists of a dehydration column, TW-802, and a small entrainer stripper column, TW-803. The entrainer is benzene and both columns operate at essentially atmospheric pressure.

The dehydration tower, TW-803, contains 50 sieve trays. The alcohol product bottoms (Stream 819) contains 99.5 percent ethanol and 0.5 percent water. This stream is cooled to 43°C in alcohol cooler HE-812, combined with the recovered fusel oils (Stream 811), and pumped to product storage (Stream 820).

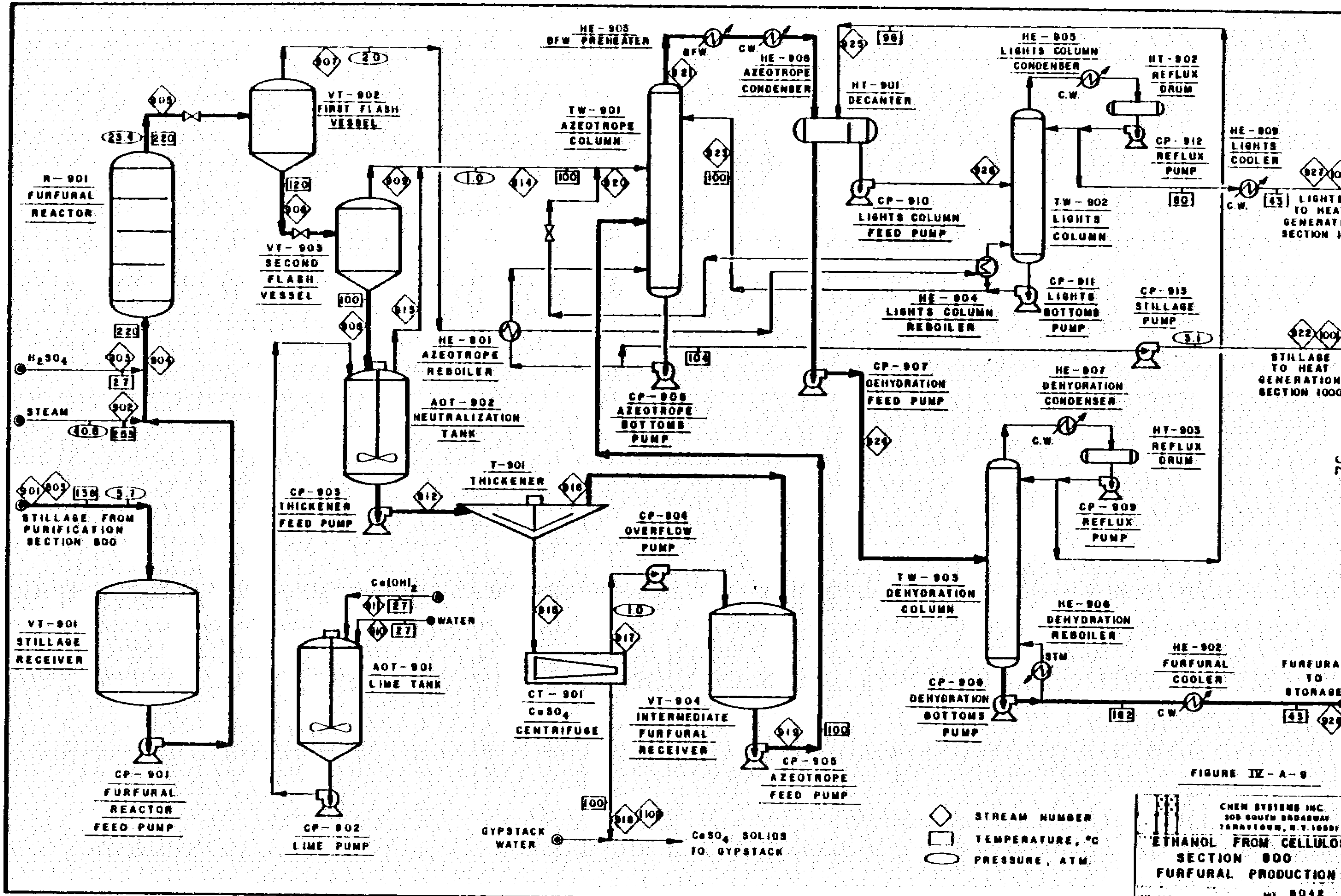
The overheads from dehydration tower TW-802 (Stream 812) is the tertiary minimum boiling azeotrope consisting of 71 percent ethanol, 17 percent water and 12 percent benzene at about 71°C. These vapors are condensed in HE-801 with the overhead vapors of entrainer stripper column TW-803 (Stream 815). These condensed vapors (Stream 816) are cooled to 64° in dehydration system overheads cooler HE-808 and enter decanter HT-801 where they separate into two layers. The upper layer (Stream 813) is hydrocarbon-rich and is pumped back to TW-802 as reflux. The lower layer (Stream 814) is the aqueous layer containing alcohol and some benzene. It is pumped to TW-803 as feed. Here, the remaining alcohol and benzene are stripped overhead and recycled to decanter HT-801. The entrainer stripper bottoms (Stream 817), essentially water, is cooled and sent to waste treatment in Section 1100. Heat is supplied to the dehydration system by condensing a portion of the overhead vapors of the beer column in reboilers HE-807 and HE-809, respectively.

9. Furfural Production (Section 900) (Figure IV-A-9)

The furfural production section can be subdivided into three subsections:

- Furfural production
- Neutralization
- Furfural recovery

Hot stillage containing approximately 5 weight percent xylose is fed from the beer still in Section 800 to stillage receiver VT-901. The stillage (Stream 901) is at 138°C and 3.7 atm. This stream is mixed with steam (Stream 902) to raise its temperature to 220°C and with sulfuric acid (Stream 903) prior to injection into furfural reactor R-901. The sulfuric acid concentration in the reactor feed is 1 weight percent.


R-901 FURFURAL REACTOR
VT-902 FIRST FLASH VESSEL
VT-903 SECOND FLASH VESSEL
HE-903 BFW PREHEATER
HE-908 AZEOTROPE CONDENSER
TW-901 AZEOTROPE COLUMN
HT-901 DECANTER
HE-905 LIGHTS COLUMN CONDENSER
HT-902 REFLUX DRUM
CP-912 REFLUX PUMP
HE-909 LIGHTS COOLER
LIGHTS TO HEAT GENERATION SECTION 1000
CP-910 LIGHTS COLUMN FEED PUMP
TW-902 LIGHTS COLUMN
CP-911 LIGHTS BOTTOMS PUMP
HE-904 LIGHTS COLUMN REBOILER
CP-913 STILLAGE PUMP
STILLAGE TO HEAT GENERATION SECTION 1000
HE-901 AZEOTROPE REBOILER
AGT-902 NEUTRALIZATION TANK
CP-906 AZEOTROPE BOTTOMS PUMP
CP-907 DEHYDRATION FEED PUMP
HE-907 DEHYDRATION CONDENSER
HT-903 REFLUX DRUM
CP-909 REFLUX PUMP
TW-903 DEHYDRATION COLUMN
HE-906 DEHYDRATION REBOILER
CP-908 DEHYDRATION BOTTOMS PUMP
HE-902 FURFURAL COOLER
FURFURAL TO STORAGE
H_2SO_4
STEAM
STILLAGE FROM PURIFICATION SECTION 800
VT-901 STILLAGE RECEIVER
CP-901 FURFURAL REACTOR FEED PUMP
CP-903 THICKENER FEED PUMP
T-901 THICKENER
CP-904 OVERFLOW PUMP
$Ca(OH)_2$
WATER
AGT-901 LIME TANK
CP-902 LIME PUMP
CT-901 $CaSO_4$ CENTRIFUGE
VT-904 INTERMEDIATE FURFURAL RECEIVER
CP-905 AZEOTROPE FEED PUMP
GYPSTACK WATER
$CaSO_4$ SOLIDS TO GYPSTACK
BFW
C.W.
STM
STREAM NUMBER
TEMPERATURE, °C
PRESSURE, ATM.
FIGURE IV-A-9
CHEM SYSTEMS INC.
ETHANOL FROM CELLULOSE
SECTION 900
FURFURAL PRODUCTION
NO 8042

The furfural reactor is a plug flow column with a 70-second residence time at 220°C. At these conditions approximately 94 percent of the xylose is converted with a 48 percent selectivity to furfural. The remainder is further converted to condensation and degradation products.

The liquid reactor effluent (Stream 905) is let down to 2 atm in first flash vessel VT-902. Flash vapors from this vessel (Stream 907) are condensed in azeotrope column reboiler HE-901 and lights column reboiler HE-904. The liquids from the first flash vessel (Stream 906) are further let down to 1 atm in second flash vessel VT-903. Flash vapors from this vessel (Stream 909) are combined with neutralization vapors (Stream 913) to form Stream 914. These vapors are combined with let down condensate from reboilers HE-901 and HE-904 and sent to the azeotrope column (Stream 920). The liquids from the second flash vessel (Stream 908) are sent to neutralization.

In neutralization tank AOT-902, calcium hydroxide is added from lime tank AOT-901 to form calcium sulfate. The heat of neutralization produces some vapors which are sent to the azeotrope column in Stream 913. The neutralized liquid proceeds to thickener T-901 where the soldis are concentrated to 20 weight percent and sent to calcium sulfate centrifuge CT-901 (Stream 915). The clarified liquid overflow (Stream 916) from T-901 is sent to intermediate furfural receiver VT-904 along with the liquid overflow from CT-901 (Stream 917). The calcium sulfate solids (Stream 918) from CT-901 at approximately 70 weight percent solids are slurried in drainage water from the gypstack and pumped to the gypstack.

Furfural solution (Stream 919) is pumped from the intermediate furfural receiver to azeotrope column TW-901. The total azeotrope column feed including Stream 920 contains approximately 1.5 weight percent furfural. This column operates at essentially atmospheric pressure and is composed of 69 sieve trays with a reflux ratio of 15.9. The furfural/water azeotrope is taken overhead along with some additional water. The overhead vapor (Stream 921), containing 99 percent of the furfural fed to recovery and composed of 23 weight percent furfural and 77 percent water, is condensed in BFW preheater HE-903 and azeotrope column condenser HE-908 at 98°C and sent to decanter HT-901 where it separates into two layers.

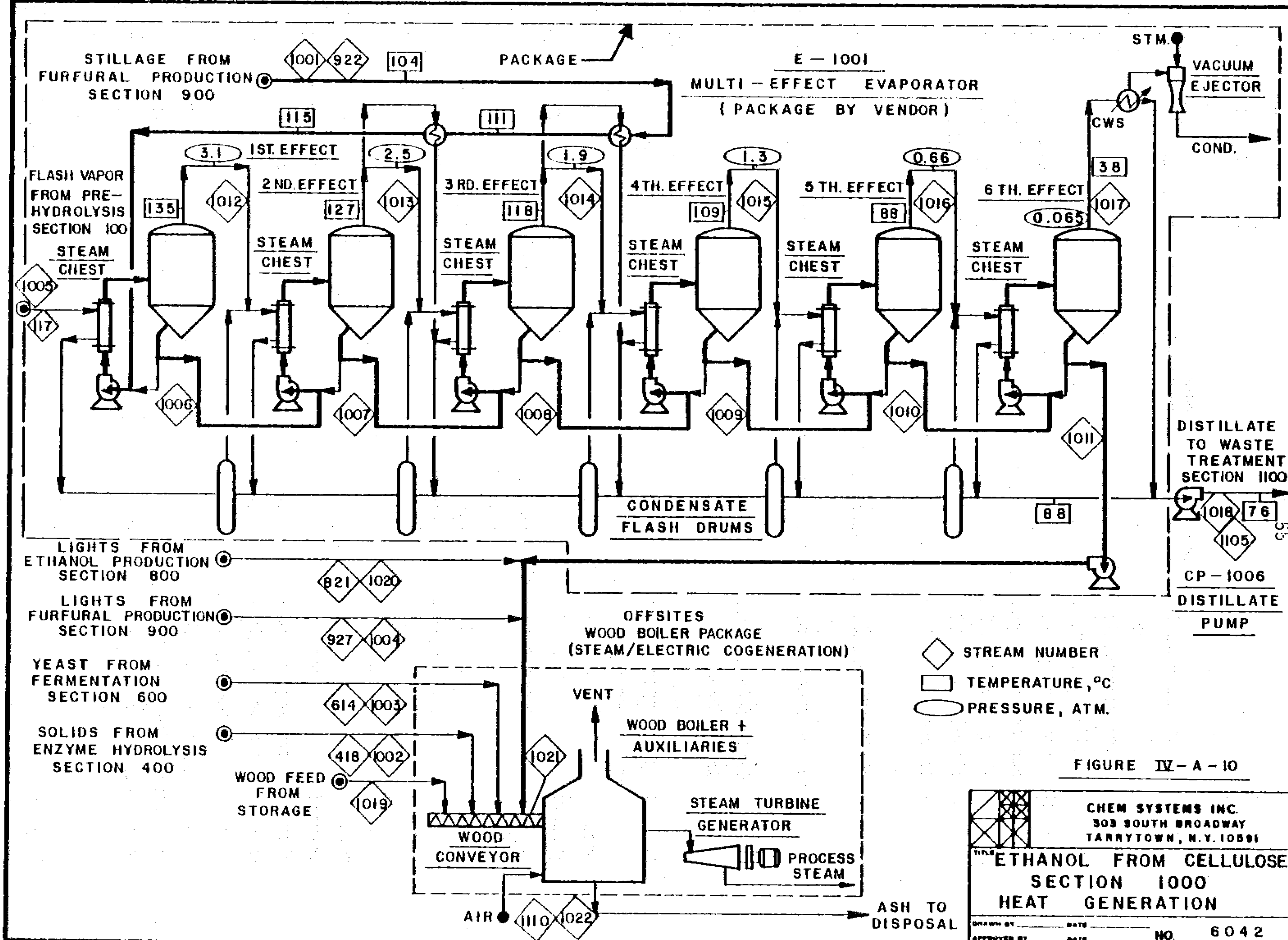
The lower organic layer composed of 84 percent furfural and 16 percent water is sent via Stream 924 to dehydration column TW-903. The upper layer composed of 18 percent furfural and 82 percent water (and possibly some lights) is sent to lights column TW-902 via Stream 926. Azeotrope column bottoms (Stream 922) is pumped to heat generation in Section 1000. Azeotrope column reboiler duty is provided via condensation of furfural reactor flash vapors in HE-901.

In lights column TW-902, any low boilers such as ethanol and methanol are removed in the overhead (Stream 927) and sent to heat generation in Section 1000. The column bottoms (Stream 923), which is essentially the aqueous layer from the decanter, is returned to the azeotrope column as reflux. This column operates at essentially atmospheric pressure.

The furfural-rich organic layer from the decanter enters dehydration column TW-903 where the remainder of the water is removed. This column operates at atmospheric pressure, has a reflux ratio of 0.34 and requires 10 sieve trays. The furfural product (Stream 928), which forms the column bottoms, is cooled to 43°C and sent to furfural storage. Its composition is 99.5 percent furfural and 0.5 percent water. The dehydration column overheads (Stream 925), which is the water/furfural azeotrope, is condensed and recycled to the decanter.

10. <u>Heat Generation (Section 1000) (Figure IV-A-10)</u>

The stillage from furfural production in Section 900 (Stream 1001) is concentrated to a 50 weight percent solution of organics in a forward feed multi-effect evaporator, E-1001. This evaporator consists of six effects operating with a forced circulation of the liquor. The system is designed with feed preheaters and condensate flash between effects to optimize the heat duty. The evaporator is a package system provided by a vendor and is constructed of carbon steel. In this system flash vapor from prehydrolysis in Section 100 (Stream 1005) is used as the heat source in the first effect.


STILLAGE FROM FURFURAL PRODUCTION SECTION 900
1001
922
104
PACKAGE
E — 1001
MULTI — EFFECT EVAPORATOR
(PACKAGE BY VENDOR)
STM.
VACUUM EJECTOR
CWS
COND.
115
111
FLASH VAPOR FROM PRE-HYDROLYSIS SECTION 100
3.1
IST. EFFECT
2.5
2 ND. EFFECT
1.9
3 RD. EFFECT
1.3
4 TH. EFFECT
0.66
5 TH. EFFECT
6 TH. EFFECT
0.065
38
135
1012
127
1013
118
1014
109
1015
88
1016
1017
STEAM CHEST
1005
117
1006
1007
1008
1009
1010
1011
CONDENSATE FLASH DRUMS
88
DISTILLATE TO WASTE TREATMENT SECTION 1100
1018
76
1105
CP — 1006
DISTILLATE PUMP
LIGHTS FROM ETHANOL PRODUCTION SECTION 800
821
1020
LIGHTS FROM FURFURAL PRODUCTION SECTION 900
927
1004
OFFSITES
WOOD BOILER PACKAGE
(STEAM/ELECTRIC COGENERATION)
YEAST FROM FERMENTATION SECTION 600
614
1003
SOLIDS FROM ENZYME HYDROLYSIS SECTION 400
418
1002
VENT
WOOD BOILER + AUXILIARIES
1021
WOOD FEED FROM STORAGE
1019
WOOD CONVEYOR
STEAM TURBINE GENERATOR
PROCESS STEAM
AIR
1110
1022
ASH TO DISPOSAL
STREAM NUMBER
TEMPERATURE, °C
PRESSURE, ATM.
FIGURE IV - A - 10
CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591
ETHANOL FROM CELLULOSE
SECTION 1000
HEAT GENERATION
NO. 6042

The distillate (Stream 1018) is sent to waste treatment in Section 1100.

The concentrated organics from the evaporator (Stream 1011) are pumped to the wood boiler system, discussed in the offsite section. In addition to the concentrated organics, there are four other waste streams that are fed to the wood boiler to supplement the main wood feed (Stream 1019). These streams are: solids from enzyme hydrolysis (Stream 1002), yeast from fermentation (Stream 1003), lights from furfural production (Stream 1004) and lights from ethanol production (Stream 1020).

11. <u>Waste Treatment (Section 1100) (Figure IV-A-11)</u>

The waste treatment section can be subdivided into two subsections:

- Condensate treatment
- Solids handling

Five plant condensate streams are combined and sent to condensate treatment system WTS-1101. These streams are flash condensate from pretreatment/prehydrolysis in Section 100 (Stream 1101), multi-effect evaporator distillate from sugar concentration in Section 500 (Stream 1102), flash condensate from preheater HE-813 in ethanol purification (Stream 1103), stripper column bottoms from ethanol purification in Section 800 (Stream 1104), and multi-effect evaporator distillate from heat generation in Section 1000 (Stream 1105).

The combined condensate (Stream 1111) contains approximately 730 pounds per hour of waste organics with a BOD of 1,560 ppm. This stream is treated in an aerobic/anaerobic system to reduce the BOD level in the clarified effluent to 30-40 ppm. After treatment, this water is reused as process water in certain sections of the plant since the suspended solids will be approximately 20 ppm.

The gypstack, WTS-1102, receives waste solids from four process neutralization steps. These are calcium sulfate solids from sugar

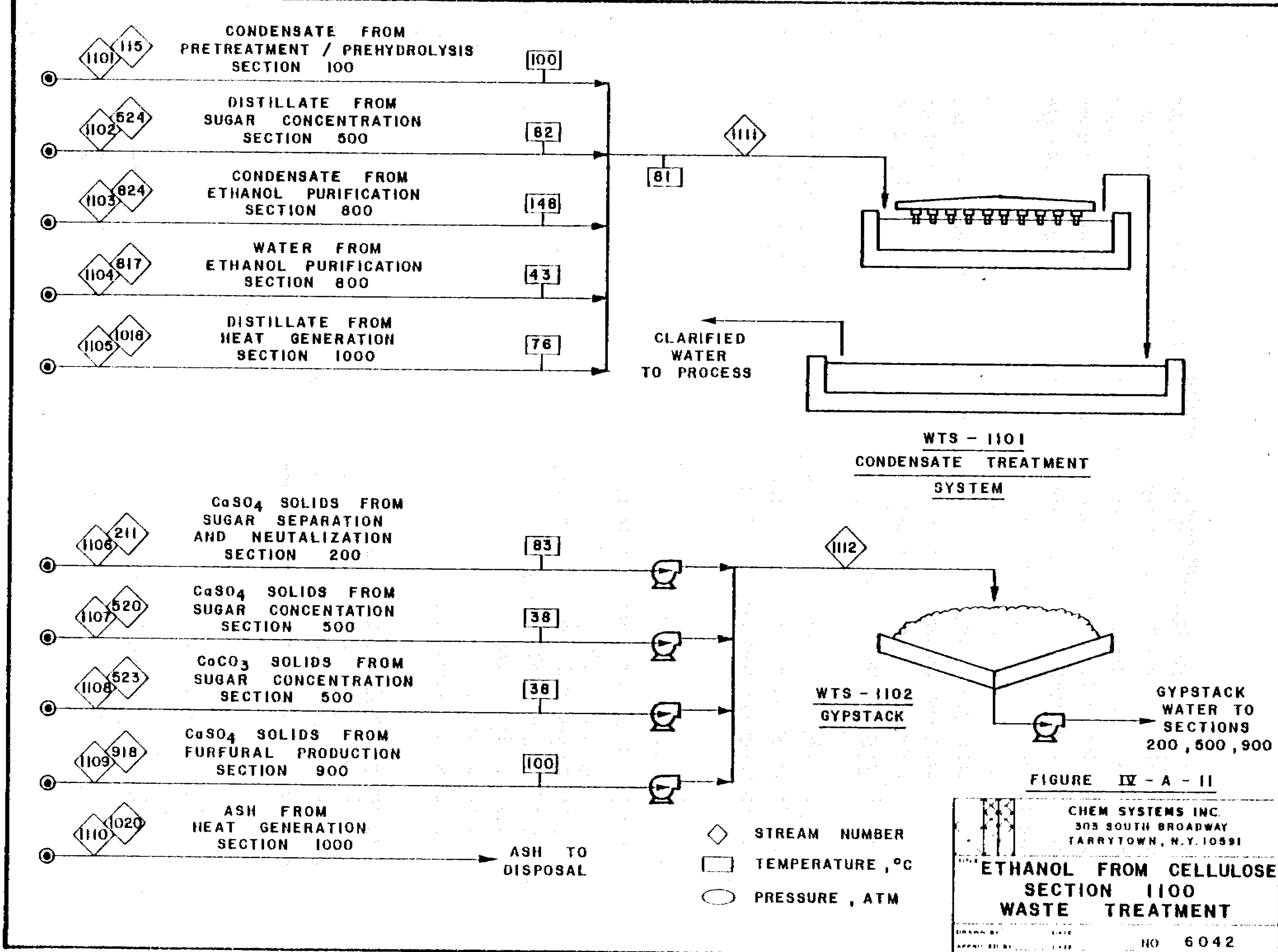


FIGURE IV - A - 11

separation and neutralization in Section 200 (Stream 1106), from sugar concentration in Section 500 (Stream 1107), and from furfural production in Section 900 (Stream 1109), and calcium carbonate solids from sugar concentration in Section 500 (Stream 1108). Each of these solids streams comes off a centrifuge and is slurried in circulating gypstack drainage water. The solids are pumped to the gypstack.

The gypstack is a lined solids storage area where the drainage liquor can be collected and pumped to the four process locations for transporting solids back to the gypstack. Approximately 6.5 tons per hour of dry solid waste is sent to the gypstack.

Ash from the heat generation steam boiler in Section 1000 (Stream 1110) can either be accumulated on the gypstack or sent directly to disposal. Approximately 1.5 tons per hour of ash are produced.

12. Offsites

The offsites in this plant consist of:

- Raw material and product storage
- Wood feed and handling
- Utility systems
- Buildings
- Waste treatment

Raw Material and Product Storage

The plant is provided with fourteen days storage for: ethanol, calcium hydroxide, sulfuric acid, caustic, corn steep liquor, furfural and sodium carbonate.

Wood Feed and Handling

The total wood requirement to the plant (as a feedstock for the process and steam boiler) is 17,500 tons per week. The design is based on the delivery of green wood chips to the plant. Based on a 5-day per week and 10-hour per day delivery schedule, the unloading facility is equpped with 2 hydraulic truck dumpers to unload 14 trucks total (25 tons per truck) per hour.

In this facility, the loaded trucks, after being weighed at the scale house, move onto the inclined dumper where the cabs are separated and the wood from the trailers is hydraulic dumped to the ground below the ramp. The chips are then moved into intermediate storage piles via front-end bulldozers. It is estimated that 20 acres are required to provide 60 days storage to allow for an inventory of wood during the bad weather months. From the piles, the wood chips are moved to a conveyor which continuously feeds a classifier where oversized chips are separated, crushed and recombined with the on-size chips. From here the chips are fed to three hold silos in parallel. From each silo the chips are metered to chip washers for the removal of dirt and sand. Finally, the washed chips are belt conveyed from each of the three washers to the refiner feed hoppers.

A separate conveyor from the classifier feeds wood to the steam boiler.

Utility Systems

Steam

The primary utility system in the plant is a steam boiler/power generator. The steam boiler is designed to produce 1,200 psia steam at 482°C which is let down to 600 and 250 psia through a steam turbine which in conjunction with a generator is used to generate electricity. The 600 and 250 psia steam is used in the process and the 250 psia steam is further let down to provide 65 psia steam where required.

The boiler system is a VU-40, CE-Power Systems type boiler fed with wood chips along with various organic waste streams as discussed in Section 1000. Based on a steam capacity of 269,000 pounds per hour of 250 psia steam and 46,100 pounds per hour of 600 psia steam, 12,400 kilowatts of power are generated. With a total plant demand of 13, 100 kilowatts, the net power import is 700 kilowatts.

Cooling Water

The cooling water system includes a cooling tower, concrete basin, circulation pumps and field erection. The system is designed for a circulation rate of 27,000 gallons per minute which is an excess of 20 percent over known plant requirements, a supply temperature of 30°C and cooling water return temperature of 44°C.

Boiler Feedwater

The boiler feedwater system is to upgrade the quality of well water to the boiler. It includes water treatment (mixed bed quality deionizer) piping, pumps, tanks and installation. The capacity of 430 gallons per minute is an excess of 30 percent over total make-up of steam used directly in the process and 2 percent of recirculated process steam.

Electrical

Although 95 percent of the power requirements in the plant are provided by cogenerated power it is assumed that the plant will be tied into the main power grid to handle 25 percent of the plant power load. The electrical system provides for a tie-in to a main utility power line, transformer, substations and main breaker, metering systems, and main power cables.

Buildings

This includes an office and administrative building, laboratory, change house and cafeteria, guard house, garage, maintenance shop and warehouse.

General Utilities

This includes fire water pumps, instrument and plant air systems (compressors, dryers and surge drums), warehouse and shop cranes.

Site Development

The site development includes site clearing and assumes no unusual problems such as environmental problems or right of ways. Included in site development are fencing, curbing, parking lots, roads, wells, drainage, rail systems, borings and general paving.

Piping

This includes piping required for instrument and plant air headers, process water, fire water loop, flare system, process area tie-ins and interconnecting piping within the storage area.

Pollution Control

Pollution control is comprised mainly of the process waste treatment pond and the gypstack discussed in Section 1100. In addition there are sanitary waste treatment systems.

B. Design Basis

A process design has been developed for a plant producing 25 million gallons per year of ethanol from mixed hardwoods. Eighty percent of the plant feedstock will be hardwoods from aspen forests and the remainder from maple/birch forests. As shown in Table IV-B-1, the average feedstock composition will be 57 percent aspen, 20 percent maple and 23 percent other hardwoods. On a moisture-free basis, the plant feed will contain approximately 47.2 percent cellulose, 31.3 percent hemicellulose and 18.5 percent lignin. The hemicellulose can be further broken down to 7.9 percent hexosans, 16.5 percent pentosans and 6.9 percent others. The moisture content is 50 percent.

The remainder of this section provides the design basis used to develop the material balances and major process equipment sizing.

TABLE IV-B-1

MIXED HARDWOOD COMPOSITION - MOISTURE FREE

	Trembling Aspen			Typical Hardwood			Red Maple			Average Plant Feedstock*		
Cellulose	51.5			43.0			40.0			47.2		
Hemicelluloses	29.6			33.1			33.9			31.3		
Hexosans		7.2			8.2			9.5			7.9	
Glucan			4.2			4.9			5.5			4.6
Mannan			2.2			2.3			3.4			2.5
Galactan			0.8			0.8			0.6			0.8
Pentosans		15.9			17.5			17.3			16.5	
Xylan			15.5			16.9			16.8			16.1
Arabinan			0.4			0.5			0.5			0.4
Other Hemis		6.5			7.5			7.1			6.9	
Uronic anhydride			3.2			3.9			3.4			3.4
Acetyl			3.3			3.6			3.7			3.5
Lignin	15.8			20.7			23.4			18.5		
Ash	0.2			0.3			0.2			0.2		
Extractives	2.9			2.9			2.5			2.8		
	100.0			100.0			100.0			100.0		

*Basis: 57% aspen, 20% maple, 23% other hardwoods.

1. Pretreatment/Prehydrolysis (Section 100)

Digester (CO-104)

Temperature	200°C
Pressure	16.0 atm
Residence time	1 minute
Maximum chip size	0.5" x 1" x 0.25"

Disk Refiner (CR-101)

Pressure	16.0 atm
Temperature	200°C
Power per dry ton per day	7 HP-day/ton

Power requirements are based on personal communications with C.E. Bauer.

Prehydrolysis Reactor (R-101)

Temperature	200°C
Pressure	15.3 atm
Residence time	12 seconds
Solids concentration	15 weight percent
Acid conc. in solubles	0.75 weight percent
Total C_6 hydrolyzed	13.6 mole percent
Glucose yield on C_6	13.4 mole percent
Total C_5 hydrolyzed	70.8 mole percent
Xylose yield on C_5	66.2 mole percent
Furfural yield on C_5	4.5 mole percent
Lignin solublized	10.0 percent
Acetic acid yield on dry wood	0.375 weight percent
Methanol yield on dry wood	0.050 weight percent

Cellulose conversions and yields are based on kinetics developed by Dartmouth[2,3]. Pentosan conversions and yields are based on kinetics developed by Dartmouth[4]. Methanol production was measured by Dartmouth[5]. Acetic acid production was assumed to be 25 percent of theoretical based on residence time in the digester and prehydrolysis reactor.

First Flash Vessel (VT-102)

Temperature	148°C
Pressure	4.4 atm

Second Flash Vessel (VT-103)

Temperature	118°C
Pressure	2.0 atm

2. Sugar Separation and Neutralization (Section 200)

Centrifuge (CT-201)

Solids recovered	95 percent
Solids concentration	35 weight percent

Rotary Drum Polishing Filter (F-201)

Solids recovered	95 percent
Solids concentration	35 weight percent

Thickener (T-201)

Solids recovered	100 percent
Solids concentration	20 weight percent

Calcium Sulfate Centrifuge (CT-202)

Solids recovered	95 percent
Solids concentration	83 weight percent

3. Enzyme Production (Section 300)

Enzyme Production Tanks (VT-301-313)

Type:	Fed batch
Temperature:	28°C
pH	5.0
Residence time	12 days
Cell density (initial)	7 grams/liter
Nutrient requirement	3 percent corn steep liquor
Cellulose concentration (initial)	1.7 percent
Total solids concentration (initial)	9.4 percent
Total solids concentration (final)	10.1 percent
Aeration rate	0.05 volume/volume/minute
Oxygen utilization	10 percent
Filter paper activity	15.0 IU/milliliter
Enzyme productivity	52 IU/liter/hour
Specific enzyme activity	0.64 IU/milligram enzyme
Carbon source requirement	0.17 grams enzyme/gram carbon source feed 109 IU/gram carbon source feed
Mycellium production	0.07 grams cells produced/gram cellulose
Cell recycle rate	55 percent

Carbon source is defined for the determination of enzyme production as the total prehydrolyzed wood feed which includes cellulose, lignin, ash, xylan, other insolubles and extractives. No credit is taken for any glucose or xylose present with respect to enzyme production and carbon source consumption. Only a fraction of the wood feed is actually utilized by the cells including cellulose, xylan and any sugars present. Complete consumption of these carbon source components is assumed. The carbon source components consumed (cellulose, xylan, xylose and glucose) plus the oxygen utilized form enzymes, cells and carbon dioxide. Carbon source consumption is independent of type of enzyme production (fed batch vs. batch) and is only dependent on carbon source type[8,9]. The carbon source, prehydrolyzed wood, produces enzymes at a carbon source consumption similar to steam exploded wood. Three percent corn steep liquor is used as the only nutrient source which satisfies all nutrient requirements and does not adversely effect enzyme production[3].

Cell Centrifuge (CT-301)

Solids recovered	95 percent
Solids concentration	15 percent

Repulping Tank (AOT-301)

Solids concentration	10 percent

Polishing Centrifuge (CT-302)

Solids recovered	95 percent
Solids concentration	15 percent
Wash rate on water and solubles	1.7 lbs/lb
Wash efficiency	95 percent

4. Enzyme Hydrolysis (Section 400)

Enzyme Hydrolysis Tanks (VT-401-406)

Type:	Continuous cascade
Temperature:	50°C
pH	4.8
Residence time	24 hours
Solids concentration (initial)	10 percent
Cellulase loading	15 IU/gram solids
Cellobiase loading	0.9 IU/gram solids
Glucose yield	90 mole percent of theoretical

Enzyme hydrolysis yields are based on data published by Dartmouth[7]. All acetic acid is neutralized with calcium hydroxide. Calcium sulfate formation via neutralization of sulfuric acid with calcium hydroxide is limited to an amount slightly below saturation at 50°C. Remaining sulfuric acid is neutralized with caustic. All xylan present is converted to xylose during enzyme hydrolysis. Cellulose not converted to glucose remains unconverted.

Centrifuge (CT-401)

Solids recovery	95 percent
Solids concentration	35 percent
Wash factor on water and solubles	0.8 lbs/lbs
Wash efficiency	88.5 percent

Polishing Filter (F-401)

Solids recovery	95 percent
Solids concentration	35 percent
Wash factor on water and solubles	0.8 lbs/lb
Wash efficiency	88.5 percent

Dewatering Press (F-402)

Solids recovery	100 percent
Solids concentration	55 percent

5. Sugar Concentration (Section 500)

Multi-Effect Evaporator (E-501)

Type:	Forward feed-forced circulation
Total evaporative capacity:	344,000 pounds H_2O/hour
Feed glucose concentration:	6.8 weight percent
Product glucose concentration:	15 weight percent

Pressures	
Steam chest first effect:	4.4 atm
Steam chest second effect:	3.1 atm
Steam chest third effect:	2.5 atm
Steam chest fourth effect:	1.9 atm
Steam chest fifth effect:	1.3 atm
Steam chest sixth effect:	0.66 atm
Vapor to condenser:	0.065 atm
Vapor temperatures	
First effect:	135°C
Second effect:	127°C
Third effect:	118°C
Fourth effect:	107°C
Fifth effect:	88°C
Sixth effect:	38°C
Steam:	148°C

Hydroclone Bottoms (Part of E-501)

Solids recovered	100 percent
Solids concentration	40 percent

$CaSO_4$ Centrifuge (CT-501)

Solids recovered	95 percent
Solids concentration	77 percent

$CaCO_3$ Centrifuge (CT-502)

Solids recovered	95 percent
Solids concentration	90 percent

6. Fermentation (Section 600)

Fermenters (VT-601-606)

Total residence time	18 hours
Temperature	30^oC
Feed glucose concentration	13.2 weight percent
Initial pH	4.0
Terminal ethanol concentration	7.0 weight percent
Glucose conversion	100 percent
Initial yeast concentration	385×10^6 cells/milliliter
Ethanol/CO_2 selectivity	93 mole percent
Glycerol/acetaldehyde/CO_2 sel.	4.9 mole percent
Fusel oil selectivity	0.1 mole percent
Yeast growth/CO_2 selectivity	2.0 mole percent
Net ethanol per pound glucose	0.475 pound
Net CO_2 per pound glucose	0.477 pound
Net glycerol per pound glucose	0.025 pound
Net acetaldehyde per pound glucose	0.012 pound
Net fusel oils per pound glucose	0.001 pound
Net yeast growth per pound glucose	0.010 pound

Yeast Centrifuge (CT-601)

Solids recovered	95 percent
Solids concentration	15 weight percent

Dewatering Press (F-603)

Solids recovered	100 percent
Solids concentration	55 weight percent

7. Carbon Dioxide Recovery (Section 700)

PACKAGE UNIT

8. Ethanol Purification (Section 800)

Beer Column (TW-801)

Feed temperature	121°C
Overhead ethanol concentration	94 weight percent
Overhead temperature	110°C
Overhead pressure	3.4 atm
Bottoms temperature	138°C
Bottoms pressure	3.7 atm

Dehydration Column (TW-802)

Entrainer	Benzene
Feed temperature	113°C
Feed ethanol concentration	92.9 weight percent
Overhead ethanol concentration	71 weight percent
Overhead benzene concentration	12 weight percent
Bottoms ethanol concentration	99.5 weight percent
Bottoms temperature	93°C
Bottoms pressure	1 atm

Entrainer Stripper Column (TW-803)

Feed temperature	64°C
Feed ethanol concentration	42 weight percent
Feed benzene concentration	7 weight percent
Overhead ethanol concentration	49.4 weight percent
Overhead benzene concentration	8.4 weight percent
Overhead temperature	85°C
Bottoms benzene concentration	0.5 weight percent
Bottoms temperature	96°C

9. Furfural Production (Section 900)

Furfural Reactor (R-901)

Temperature	220°C
Pressure	23.4 atm
Residence time	70 seconds
Acid concentration in solubles	1.0 weight percent
Xylose feed concentration	4.8 weight percent
Xylose converted	92.3 percent
Selectivity to furfural	51.6 mole weight

Xylose conversions are based on kinetics developed by Dartmouth[4].

First Flash Vessel (VT-902)

Temperature	120°C
Pressure	2.0 atm

Second Flash Vessel (VT-903)

Temperature	100°C
Pressure	1.1 atm

Thickener (T-901)

Solids recovered	100 percent
Solids concentration	20 percent

Calcium Sulfate Centrifuge (CT-901)

Solids recovered	95 percent
Solids concentration	61 weight percent

Azeotrope Column (TW-901)

Feed furfural concentration	1.5 weight percent
Feed temperature	100°C
Reflux ratio	15.9
Actual trays	69
Overhead temperature	98°C
Overhead pressure	1.0 atm
Overhead furfural concentration	23 weight percent
Bottoms temperature	104°C
Furfural recovery in overheads	99 percent

Design parameters are based on column simulations developed by Smuk[10].

Lights Column (TW-902)

Feed lights concentration	0.06 weight percent
Feed temperature	98°C
Actual trays	15

Dehydration Column (TW-902)

Feed furfural concentration	84 weight percent
Feed temperature	98°C
Reflux ratio	0.34
Actual trays	10
Overhead temperature	98°C
Overhead pressure	1.0 atm
Overhead furfural concentration	35 weight percent
Bottoms temperature	162°C
Bottoms furfural concentration	99.5 weight percent

10. Heat Generation (Section 1000)

Multi-Effect Evaporator (E-1001)

Type:	Forward feed-forced circulation
Total evaporative capacity:	327,224 pound H_2/hour
Pressures	
Steam chest first effect:	4.4 atm
Steam chest second effect:	3.1 atm
Steam chest third effect:	2.5 atm
Steam chest fourth effect:	1.9 atm
Steam chest fifth effect:	1.3 atm
Steam chest sixth effect:	0.66 atm
Vapor to condenser:	0.065 atm
Vapor temperature	
First effect:	135°C
Second effect:	127°C
Third effect:	118°C
Fourth effect:	109°C
Fifth effect:	88°C
Sixth effect:	38°C
Steam:	148°C

11. Waste Treatment (Section 1100)

Condensate Treatment System (WTS-1101)

Inlet suspended solids	1,000 PPM
BOD loading	1,560 PPM
Effluent suspended solids	20 PPM

Based on personal communications with EIMCO.

C. Material Balance

A material balance for a plant producing 25 million gallons per year of ethanol from mixed hardwoods has been developed. The detailed material balance is presented in Appendix A by plant section. Overall raw material requirements are summarized in Table IV-C-1 and Table IV-C-2 summarizes plant product and waste effluent streams.

TABLE IV-C-1

PROCESS RAW MATERIALS

	Pounds/Hour
Wood chips at 50 percent moisture	
To process - Stream 101	185,812
To boiler - Stream 1019	23,155
	208,967
Sulfuric acid - anhydrous	
Stream 105	3,847
Stream 604	14
Stream 903	3,112
	6,973
Calcium hydroxide - anhydrous	
Stream 206	2,477
Stream 403	350
Stream 911	2,366
	5,193
Sodium hydroxide - anhydrous	
Stream 303	139
Stream 403	634
	773
Sodium carbonate - anhydrous	
Stream 502	378
Corn steep liquor at 46 percent moisture	
Stream 305	1,473
Cellobiase at 65 percent moisture	
Stream 406	241
Benzene	
Stream 822	7

The plant requires 2,508 tons per day of green hardwood chips. Approximately 89 percent of this feedstock is used in the process and the remainder is steam boiler fuel. Ethanol production is 75,000 gallons per day. This is a fuel grade product containing 0.5 weight percent water and 0.2 weight percent fusel oils. Two major by-products are also recovered, carbon dioxide and furfural. The carbon dioxide production rate is 244 tons per day. Approxiamately 55 tons per day of furfural are produced at a purity of 99.5 weight percent.

TABLE IV-C-2

PRODUCT, BY-PRODUCTS AND WASTE STREAMS

	Pounds/Hour
Ethanol at 0.5 percent moisture Stream 820	20,576
Carbon dioxide - anhydrous Stream 703	20,350
Furfural at 0.5 percent moisture Stream 928	4,562
Waste solids - anhydrous basis Stream 1112	12,580
Condensed organics - anhydrous basis Stream 1111	727
Ash Stream 1110	3,056

D. Process Utility Summaries

1. Heat Balance Summary

The overall plant heat balance is summarized in Table IV-D-1 in terms of major process thermal interchanges. This table excludes heat exchangers that are located entirely within package units. Also, heat loads for sterlization, ejector operation and steam cleaning are not included in this table although they have been included in the utility summary.

TABLE IV-D-1

HEAT BALANCE SUMMARY

Item	Stream Heated	Stream Cooled	MM BTU/HR
CO-104	Wood Chip Feed	Steam	40.7
R-101	Prehydrolysis Feed	Steam	53.5
HE-101	Water to Prehydrolysis	Flash Vapors	17.3
HE-102	Water to Prehydrolysis	Flash Vapors	26.2
HE-201	Neutralized Solids	Centrifuge Overflow	9.4
HE-202	Cooling Water	Centrifuge Overflow	12.4
HE-301-314	Well Water	Enzyme Production	2.9
HE-407	Cooling Water	Enzyme Hydrol. Feed	24.4
E-501	Sugar Solution	Steam	68.6
E-501	Cooling Water	Evaporator Vapors	82.3
HE-607	Cooling Water	Sugar Solution	3.5
HE-601-606	Well Water	Fermentation	10.9
HE-801	Beer Feed	Dehydration Overheads	16.1
HE-802	Beer Feed	Beer Col. Overheads	10.2
HE-813	Beer Feed	Prehydrol. Flash Vapors	9.2
HE-803	Beer Feed	Steam	7.8
HE-804	Beer Col. Reboiler	Steam	52.4
HE-806	Cooling Water	Beer Column Vent	0.5
HE-807	Dehyd. Column Reboiler	Beer Col. Overheads	29.4
HE-808	Cooling Water	Dehydration Overheads	13.4
HE-809	Stripper Reboiler	Beer Col. Overheads	4.8
HE-810	Cooling Water	Fusel Oil	0.5
HE-811	Cooling Water	Waste Water	0.1
HE-812	Cooling Water	Ethanol Product	0.6
R-901	Furfural Reactor Feed	Steam	37.1
HE-901	Azeotrope Reboiler	Flash Vapors	50.8
HE-903	Boiler Feedwater	Azeotrope Col. Overheads	16.6
HE-908	Cooling Water	Azeotrope Col. Overheads	48.7
HE-902	Cooling Water	Furfural Product	1.0
HE-904	Lights Col. Reboiler	Flash Vapors	9.0
HE-905	Cooling Water	Lights Col. Overheads	9.0
HE-906	Dehydration Reboiler	Steam	1.2
HE-907	Cooling Water	Dehydrated Overheads	1.2
HE-909	Cooling Water	Lights Waste Stream	0.0
E-1001	Stillage	Prehydrol. Flash Vapors	44.3
E-1001	Cooling Water	Evaporator Vapors	67.8

2. Steam Requirements

Process steam requirements are listed in Table IV-D-2. Three pressure levels (600 psia, 250 psia and 65 psia) are generated via letdown of 1,200 psia, 482°C steam through power generation turbines. The total steam required is 315,100 pounds per hour.

TABLE IV-D-2

UTILITY SUMMARY - STEAM
(Pounds per Hour)

Item	Name	65 psia	250 psia	600 psia
CO-104	Digester	-	48,609[1]	-
R-101	Prehydrolysis reactor	-	63,932[1]	-
Subtotal Section 100		-	112,541	-
F-302	Filter	2,400[1]	-	-
Subtotal Section 300		2,400	-	-
VT-401-406	Enzyme tanks	500	-	-
Subtotal Section 400		500	-	-
E-501	Multi-effect evaporator	77,000	-	-
E-501	Ejector	500[1]	-	-
Subtotal Section 500		77,500	-	-
F-602	Filter	500[1]	-	-
Subtotal Section 600		500	-	-
HE-803	Final beer preheater	3,590	-	-
HE-804	Beer column reboiler	-	63,500	-
Subtotal Section 800		3,590	63,500	-
R-901	Furfural reactor	-	-	46,131[1]
HE-906	Dehydration reboiler	-	1,450	-
Subtotal Section 900		-	1,450	46,131
E-1001	Ejector	500[1]	-	-
Subtotal Section 1000		500	-	-
Steam cleaning		1,500[1]	-	-
Subtotal miscellaneous		1,500	-	-
Total process steam		5,400	112,541	46,131
Total exchanger steam		86,090	64,950	-
Plant total		91,490	177,491	46,131
Steam generated		91,490	177,491	46,131

[1]Process steam

3. Cooling Water Requirements

Cooling water requirements are summarized in Table IV-D-3. On the basis of a 14°C temperature rise, the total cooling water load is 22,700 gallons per minute. The cooling water supply is available at 30°C. Since approximately five percent of the cooling capacity is required at a temperature below 30°C, fresh well water is assumed for this supply which also acts as the make-up for the cooling water system.

TABLE IV-D-3

UTILITY SUMMARY - COOLING WATER
(14°C Temperature Rise)

Item	Name	GPM
HE-202	Filter Cooler	991
Subtotal	Section 200	991
HE-301-314	Coolers	232
Subtotal	Section 300	232
HE-407	Slurry coolers	1,950
Subtotal	Section 400	1,950
E-501	Evaporator condenser	6,579
Subtotal	Section 500	6,579
HE-601-606	Coolers	871
HE-607	Sugar cooler	280
Subtotal	Section 600	1,151
HE-805	Trim condenser	-
HE-806	Vent condenser	40
HE-808	Dehydration overhead cooler	1,470
HE-810	Fusel oil cooler	40
HE-811	Waste water cooler	8
HE-812	Alcohol cooler	48
Subtotal	Section 800	1,606
HE-902	Furfural cooler	80
HE-905	Lights column condenser	719
HE-907	Dehydration condenser	96
HE-908	Azeotrope condenser	3,893
HE-909	Lights cooler	2
Subtotal	Section 900	4,790
E-1001	Evaporator condenser	5,420
Subtotal	Section 1000	5,420
Plant total		22,719

4. Process Water Requirements

Process water requirements are summarized in Table IV-D-4. The total process water load of 1,030 gallons per minute can be supplied by treated water from the consensate treatment system in Section 1100.

TABLE IV-D-4

UTILITY SUMMARY - PROCESS WATER

Item	Name	GPM
R-101	Prehydrolysis reactor	607
Subtotal Section 100		607
AOT-201	Lime tank	39
Subtotal Section 200		39
VT-301-314	Enzyme production tanks	45
VT-315	Wash water tank	156
Subtotal Section 300		201
AOT-401	Slurry tank	53
VT-407	Wash water tank	77
Subtotal Section 400		130
TW-601	Scrubber	44
Subtotal Section 600		44
DDT-801	Fusel oil washer/decanter	2
Subtotal Section 800		2
AOT-901	Lime tank	5
Subtotal Section 900		5
Plant total process water		1,028
Water from condensate treatment system		1,160
Net import process water		0

5. Boiler Feedwater Requirements

Boiler feedwater is required to make up for the steam injected into process as well as the two percent of circulating exchanger steam assumed as losses. The total boiler feedwater make-up is 334 gallons per minute as summarized in Table IV-D-5.

TABLE IV-D-5

UTILITY SUMMARY - BOILER FEEDWATER

Item	Name	GPM
CO-104	Digester	97
R-101	Prehydrolysis reactor	128
Subtotal Section 100		225
F-302	Filter	5
Subtotal Section 300		5
E-501	Ejector	1
Subtotal Section 500		1
F-602	Filter	1
Subtotal Section 600		1
R-901	Furfural reactor	92
Subtotal Section 900		92
E-1001	Ejector	1
Subtotal Section 1000		1
Miscellaneous		3
BFW make-up		6
Subtotal miscellaneous		9
Plant total boiler feedwater		334

6. Power Requirements

Horsepower requirements are presented in Table IV-D-6. This tabulation is based on the design specifications for all plant equipment items and results in a total design horsepower load of 19,960. A number of major power users will actually be operating at less than design capacity. Thus, the steady-state power load is 13,100 kilowatts. Power cogeneration from high pressure steam (Section IV-A-12) yields 12,400 kilowatts. The net imported power from the utility grid is thus 700 kilowatts.

TABLE IV-D-6

UTILITY SUMMARY - POWER

Item	Name	Design HP
CR-101	Disk Refiner	9,000
CO-103	Feeder (in CR-101)	-
CO-104	Digester (in CR-101)	-
R-101	Prehydrolysis reactor	5
CP-101	Water feed pump	150
Subtotal Section 100		9,155
CT-201	Centrifuge	50
CP-201	Overflow pump	30
F-201	Rotary drum polishing filter	3
AOT-201	Lime tank	30
AOT-202	Neutralization tank	40
CP-202	Lime pump	2
CP-203	Thickener feed pump	25
CO-201	Conveyor	40
CO-202	Conveyor	7.5
T-201	Thickener	10
CT-202	Centrifuge	25
CP-204	Overflow pump	3
CP-205	Sugar transfer pump	25
Subtotal Section 200		290.5
CO-301	Conveyor	7.5
CP-301-314	Recirculation pumps	260
FN-301	Air blower	10
CP-315-328	Discharge pumps	260
CP-329	Cell centrifuge feed pump	20
CT-301	Cell centrifuge	25
AOT-301	Repulping tank	5
CP-330	Wash water pump	10
CT-302	Polishing centrifuge	150
CP-331	Centrifuge feed pump	5
CP-332	Cell recylce pump	5
CP-333	Enzyme feed pump	7.5
Subtotal Section 300		765
AOT-401	Slurry tank	50
CP-413	Feed pump	60
A-401-406	Agitators	375
CP-407-412	Discharge pumps	500
CT-401	Centrifuge	750
CP-415	Overflow pump	60
F-401	Rotary drum polishing filter	3
CP-414	Wash water pump	5
CO-401	Conveyor	15
F-402	Dewatering press	225

TABLE IV-D-6
(Continued)

UTILITY SUMMARY - POWER

Item	Name	Design HP
CP-416	Press discharge pump	5
CP-417	Sugar transfer pump	50
Subtotal Section 400		2,098
E-501	Multi-effect evaporator	1,200
CT-501	Calcium sulfate centrifuge	1
CP-501	Sugar pump	20
CP-502	Overflow pump	0.5
AOT-501	Precipitator	4.0
AOT-502	Sodium carbonate solution tank	0.5
CP-503	Sodium sulfate pump	0.5
CT-502	Calcium carbonate centrifuge	250
CP-504	Overflow pump	25
CP-505	Filter feed pump	25
CP-506	Distillate pump	25
Subtotal Section 500		1,587.5
FN-601	Blower	75
FN-602	Air blower	10
CP-613	Fermentation feed pump	25
CP-601-606	Recirculation pumps	125
CP-607-612	Discharge pumps	125
CP-613	Scrubber return pump	2
CT-601	Yeast centrifuges	400
CP-616	Overflow pumps	30
Cp-615	Underflow pumps	6
CP-617	Yeast pump	3
F-603	Dewatering press	1
CP-618	Discharge pump	1
AOT-601	Yeast hold tank	40
CO-601	Conveyor	7.5
CP-619	Purification feed pump	40
Subtotal Section 600		890.5
Subtotal Section 700		2,100
CP-804	Dehydration reflux pump	5
CP-805	Entrainer/stripper feed pump	2
CP-806	Entrainer column bottoms pump	2
CP-801	Beer column bottoms pump	25
CP-802	Reflux pump	20
CP-807	Wash water pump	0.5
CP-808	Fusel oil pump	0.5
CP-803	Dehydration column bottoms pump	2
Subtotal Section 800		57

TABLE IV-D-6
(Continued)

UTILITY SUMMARY - POWER

Item	Name	Design HP
CP-901	Furfural reactor feed pump	40
AOT-902	Neutralization Tank	25
CP-902	Lime pump	1
AOT-901	Lime tank	20
CP-903	Thickener feed pump	20
T-901	Thickener	5
CT-901	Calcium sulfate centrifuge	25
CP-904	Overflow pump	2
CP-905	Azeotrope feed pump	20
CP-910	Lights column feed pump	7.5
CP-907	Dehydration feed pump	1
CP-912	Reflux pump	0.5
CP-911	Lights bottoms pump	7.5
CP-906	Azeotrope bottoms pump	25
CP-913	Stillage pump	25
CP-909	Reflux pump	0.5
CP-908	Dehydration bottoms pump	2
Subtotal Section 900		227
CP-1006	Distillate pump	25
E-1001	Multi-effect evaporator	1,100
Subtotal Section 1000		1,125
WTS-1101	Condensate treatment system	600
WTS-1102	Gypstack	60
Subtotal Section 1100		660
	Wood handling	
	Steam boiler	
	Cooling water system	
	Miscellaneous	
Subtotal offsites		1,000
Total plant design horsepower		19,955.5
Actual running horsepower		17,580
Total plant kilowatts		13,100
Generated power, kilowatts		12,400
Net import power, kilowatts		700

E. Plot Plan

The 25 million gallon per year ethanol facility is located in Gladwin County, Michigan. If centrally located within the county, the plant site would be midway (approximately 15 miles) from Interstate 75, U.S. 27 and U.S. 10, thus assuring reasonable highway access to all markets within the state and to the sources of wood feedstock. The facility itself requires a 130-acre site with a major portion of space reserved for wood storage and the gypstack (approximately 50 acres). The main facilities including the battery limits plant, powerhouse and waste treatment occupy 25 acres. Storage (other than wood) is sited on 10 acres and the administration facilities including a truck staging area occupy 20 acres. A preliminary plot plan for this facility is shown in Figure IV-E-1.

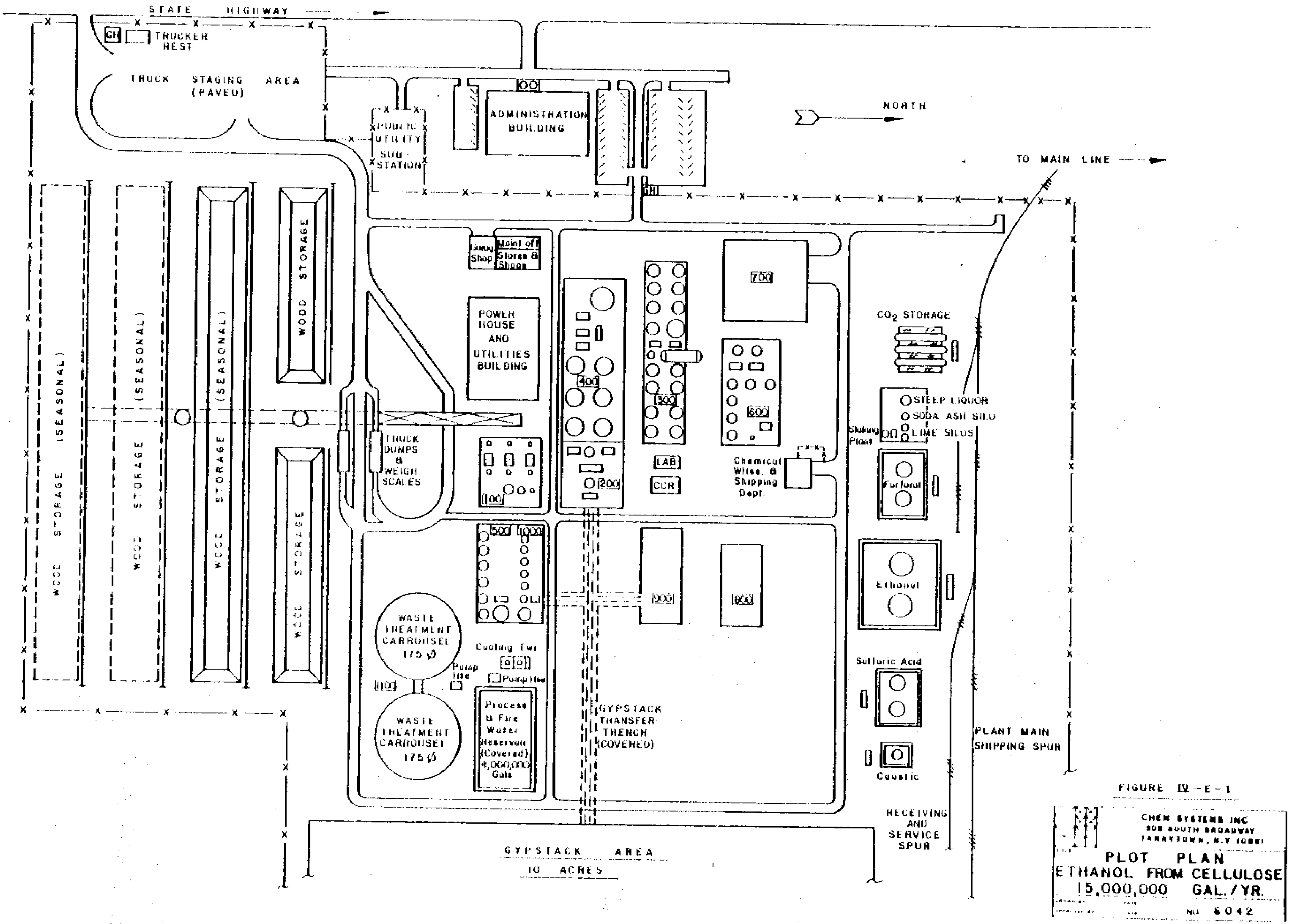


FIGURE IV-E-1

PLOT PLAN
ETHANOL FROM CELLULOSE
15,000,000 GAL./YR.

V. SITING CONSIDERATIONS

A. Selection of Michigan Site

1. Introduction

The selection of a Midwest site for the ethanol facility and the selection of the wood feedstock based on available forest types are critical to the economic success of a project of this type. As illustrated in Table III-A-2, Aspen is an ideal hardwood for this facility based on potential reducing sugars. This results in high ethanol yields per unit of cost of wood feedstock. Thus, site selection was based on screening locations with abundant aspen resources. The selection process involved screening ethanol consumption patterns, gasoline demand, applicable tax laws, wood resource acreage and cost elements for delivered wood.

Based on these issues, Michigan was determined to be an ideal state in which to locate the proposed ethanol plant. Gladwin County, centrally located within Michigan's northern lower peninsula, was chosen as the plant site because of its access to abundant aspen forests as well as ethanol markets within the state.

In the following subsections, some of the major issues which affected the selection of the plant site and feedstock will be discussed in detail.

2. Feedstock Availability and Pricing

The state of Michigan contains 17.5 million acres of commercial forest land. As shown in Table V-A-1, twenty percent of the commercial forest land is aspen forest type. Another 35 percent is maple-birch type which also can be used in the ethanol facility although its ethanol yield will be somewhat lower. More than half of the state aspen forests can be found in the northern lower peninsula.

TABLE V-A-1

MICHIGAN HARDWOOD RESOURCES
(Thousand Acres)

	Northern Lower Peninsula	Southern Lower Peninsula	Upper Peninsula	Total
Total forest land	6,930	2,549	8,891	18,370
Commercial	6,695	2,463	8,332	17,490
Aspen	1,818	184	1,404	3,406
% aspen of comm'l	27	8	17	20
Maple-birch	1,662	903	3,532	6,097
% M.B. of comm'l	25	37	42	35

The aspen forest distribution by county within the northern lower peninsula is shown in Table V-A-2. It can be seen that Gladwin County contains 7 percent of the total state aspen forest land. When coupled with the four surrounding counties of Clare, Midland, Ogemaw and Roscommon, the aspen forest resource base constitutes 24 percent of the state total or 820,000 acres. Approximately 550,000 acres or 9 percent of the total state maple-birch forest land are also contained in these five counties. A map illustrating the aspen distribution throughout the Michigan lower peninsula is presented in Figure V-A-1. This map also shows the central location of Galdwin county.

The species of trees found in aspen and maple-birch forests in the northern lower peninsula are listed in Table V-A-3. Approximately 56 percent of the aspen forest is aspen and 45 percent of the maple-birch forest is maple-birch. With the exception of the softwoods, all the other species could be used as feedstock in the ethanol facility. If it is assumed that 30 percent of the plant feedstock will be hardwoods from aspen forests and the remainder from maple-birch forests, the average feedstock composition will be 57 percent aspen, 20 percent maple and 23 percent other hardwoods.

TABLE V-A-2

ASPEN DISTRIBUTION BY COUNTY
(% of Total State Forest Type)

	Aspen	Maple-Birch	Percent of County Total Forest Land
Alcona	6	2	45
Alpena	6	1	59
Cheboygan	6	7	61
Clare	4*	2	52
Gladwin	7	1	73
Midland	4*	2	66
Montmorency	5	5	58
Ogemaw	5*	2	52
Oscoda	5	2	38
Presque Isle	5	2	45
Roscommon	4*	2	37
Total	57	28	

*Within reasonable sourcing area of Gladwin County/Midland, Bay City site.

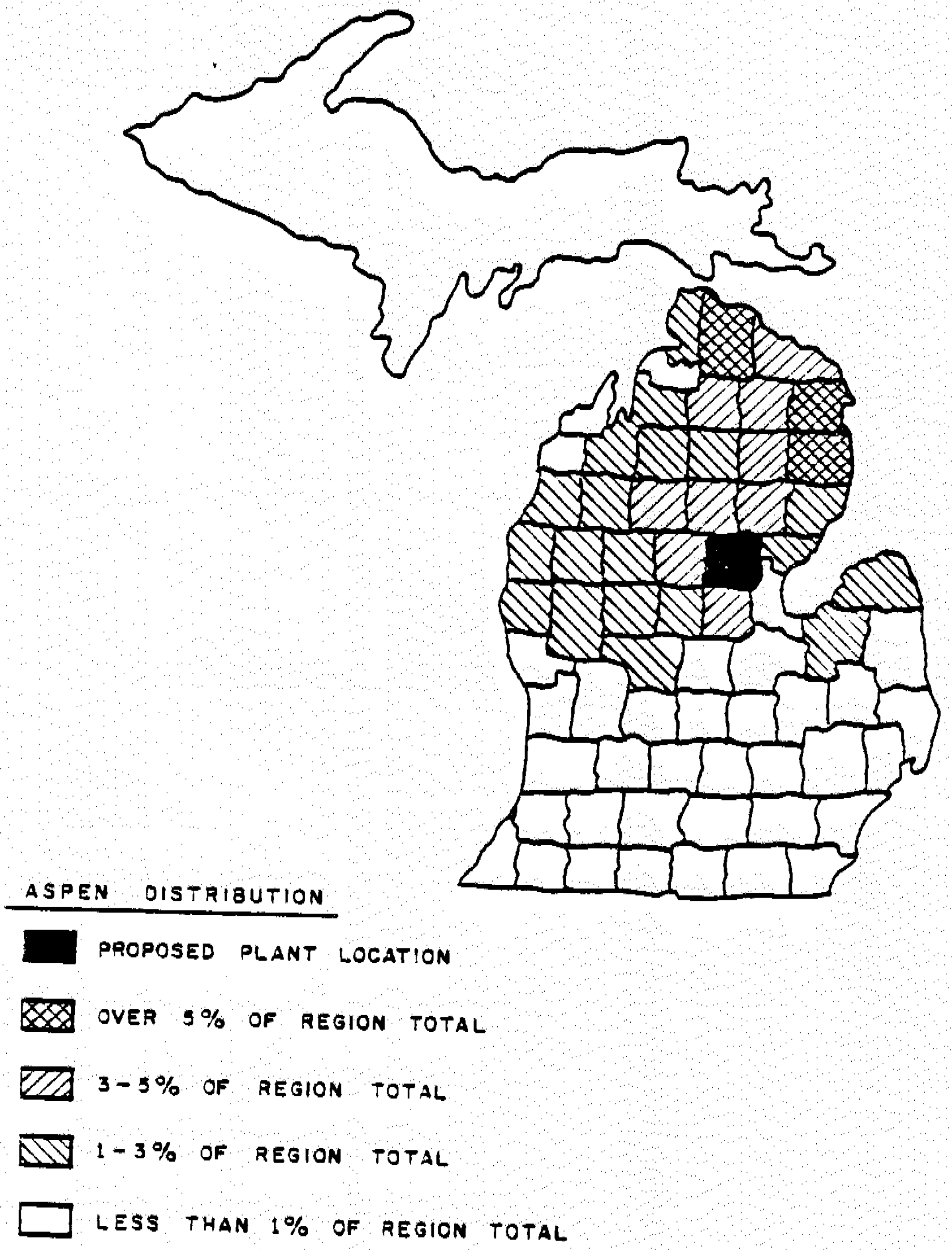
TABLE V-A-3

SPECIES OF TREES IN THE NORTHERN LOWER PENINSULA
(Percent of Total)

	Aspen forest Type	Maple-Birch Forest Type
Big Tooth Aspen	28	5
Quaking Aspen	28	5
Other hardwoods		
Maple-Birch	12	45
Red Oaks	7	5
Paper Birch	5	3
Balsam Poplar	4	-
Ash	2	6
Beech	-	5
Others	2	13
Softwoods	12	7

FIGURE V-A-1

MICHIGAN LOWER PENINSULA
ASPEN RESOURCES



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PROJECT NO. 6042 DATE

The ownership of forest land in the northern lower peninsula is shown in Table V-A-4. This table also illustrates the average age and growth rate for the two forest types. It can be seen that the aspen forests are younger and have a higher growth rate. Based on feedstock requirements of 104 tons per hour or 33 million cubic feet per year of wood, Gladwin County could supply more than half these needs simply from the average annual growth rate. Gladwin and its four neighboring counties combined grow enough wood to keep two facilities of this size going.

TABLE V-A-4

NORTHERN LOWER PENINSULA

HARDWOOD RESOURCES

	Aspen	Maple-Birch
Acreage, 000	1,818	1,662
Ownership, %		
National forest	9	4
State	29	24
County, municipal	1	-
Farmer	16	30
Misc. pvt. corp.	8	6
Misc. pvt. indiv.	35	35
Forest ind.	1	1
Miscellaneous	-	-
Total	100	100

Average Growth/Year (Cu ft/acre)	% of Aspen	% All Types	% of Maple-Birch	% of All Types
85+	44	47	24	24
20-84	56	21	76	25
0-21	0	32	0	51

Age Class	Aspen	Maple-Birch
1-2 yrs.	44	5
21-50	24	27
50+	32	68

If all the wood came from these five counties, the average hauling distance would be 30-40 miles. A more conservative 50-mile hauling distance has been used in this study. Feedstock pricing is based on whole tree chipping performed in the forest by contract crews who supply their own equipment. Currently, hauling costs are two dollars per loaded mile for 25-ton loads. With an assumed average distance of 50 miles. The plant gate cost for hardwood chips becomes 18 dollars per wet ton as shown in Table V-A-5.

TABLE V-A-5

WOOD FEEDSTOCK COST - MID 1984
(Dollars per Wet Ton)

Stumpage	3.0
Harvesting and chipping	11.0
Hauling	4.0
Total	18.0

Three different hardwood pricing scenarios were formulated for the plant operating years, 1987-2001:

- Escalation at the GNP deflator.
- Escalation to reach equivalency with softwood prices by the year 2000 with softwoods escalating at the GNP deflator.
- Escalation according to Chem Systems projections for Midwest coal prices.

Chem Systems projections for the GNP deflator and Midwest coal price escalation are given in Table V-A-6. The base case ethanol facility economics assume that the hardwood chips will escalate with the GNP deflator. The second scenario is based on the assumption that demand for hardwoods within the pulp and paper industry and the lumber industry will increase in Michigan as it finds increasing use in oriented strand board, pulp and paper, etc. The hardwood demand would equal that for softwoods by the year 2000 if these industries continue to increase their usage of hardwoods. Currently, hardwood is priced at approximately 88 percent of softwoods because of its lower demand by the pulp and paper industry. The third scenario assumes that a major use for hardwood in Michigan is as a fuel. If this were the dominant future use for hardwood, its competition would be coal, and therefore its value would escalate with Midwest coal prices. These pricing scenarios are summarized in Table V-A-7.

TABLE V-A-6

ESCALATION FACTORS

	GNP Deflator (%)	Midwest Coal Price Escalation (%)
1984-1990	5.0	8.0
1991-2001	4.5	7.0

TABLE V-A-7

PLANT GATE

HARDWOOD PRICE PROJECTIONS
Dollars per Wet Ton)

	Softwoods @ GNP Deflator	Hardwoods @ GNP Deflator	Hardwoods Via Coal Escalation	Hardwoods Approaching Softwoods
1984	20.5	18.0	18.0	18.0
1987	23.7	20.8	22.7	21.3
1990	27.5	24.1	28.6	25.4
1995	34.2	30.1	40.1	32.9
2000	42.7	37.5	56.2	42.7
2001	44.6	39.1	60.1	44.6

3. Gasoline Pool Analysis

The composition of demand for refined products in the United States has and will continue to be distinctly different from other industrialized regions of the world. Traditionally high in gasoline and low in residual fuel oil demand, consumption in the United States has resulted in a refining industry that is a highly complex structure designed to produce and market motor gasoline. The decline in gasoline demand, predicted to continue through the 1990s, is a major factor in the planning and actual operation of the petroleum refining industry. Similarly, demand for residual fuel oil will continue its recent decline and, again, refiners must adjust their operations accordingly. Kerosene/jet fuel and distillate fuel will experience a rise in demand level and will thus represent an increasing share of the composite refined product demand slate as shown in Table V-A-8, which presents the demand for the key products.

TABLE V-A-8

U.S. MAJOR REFINED PRODUCT DEMAND, 1980-1990
(MBPCD)

Product	1980	1982	1983	1985	1990	Annual Percent Change 1980-1990
Gasoline	6,579	6,539	6,617	6,600	6,000	(0.9)
Kerosene/kerojet	980	919	935	970	1,015	0.4
Distillate fuel	2,866	2,671	2,682	2,870	3,050	0.6
Residual fuel	2,508	1,716	1,403	1,972	1,805	(3.2)
Total	12,933	11,845	11,637	12,412	11,870	(0.9)

Gasoline Demand

Since fermentation ethanol is largely consumed for transport fuel applications both as an octane enhancer and volume extender, the emphasis in this section will be on current and anticipated demand for gasoline and how the dynamics of this market will affect ethanol consumption during the 1980s. In the United States, transport fuels account for over 60 percent of refined product demand, with gasoline being the single largest volume product consumed. Although overall gasoline use is projected to decline, it will still account for about 40-45 percent of overall petroleum product demand in 1990, about the same share that occurred in 1978, the peak year for U.S. gasoline and refined product consumption.

Demand by Grade

Although total gasoline demand is expected to drop over the study period, the demand for unleaded grades will increase. The addition of lead compounds to improve the octane ratings of motor gasoline has been severely restricted by environmental regulations. Lead phasedown has caused the clear octane (octane prior to the addition of lead) of the U.S. gasoline pool to rise significantly and will cause the clear pool to increase further. Consistent with increasing demands for unleaded grades, in general, are increases in the demand for premium unleaded gasoline. Of the total gasoline demand, unleaded premium grades will increase from about 12 percent in 1983 to about 27 percent in 1990. The octane pool increase is displayed in graph form in Figure V-A-2. It is this change in the clear octane pool that provides the impetus for using octane enhancers, reflecting a growing requirement for ethanol and other octane enhancers.

Ethanol demand has grown from less than 100 million gallons in the 1979-1981 period to about 225 million gallons in 1982 and just over 400 million gallons in 1983. Examining the projected demand for gasoline by grade, U.S. refining capacity, the use of alternative octane enhancers, the status of state tax exemption laws, as well as current and anticipated ability to produce anhydrous fermentation grade product, it is estimated that overall U.S. ethanol consumption for transport fuel applications will equal about 600 million gallons in 1985 and grow to about 350 million gallons by 1990, as shown in Table V-A-9.

FIGURE V-A-2

AVERAGE U.S. GASOLINE POOL OCTANE
(R+M)/2 - CLEAR

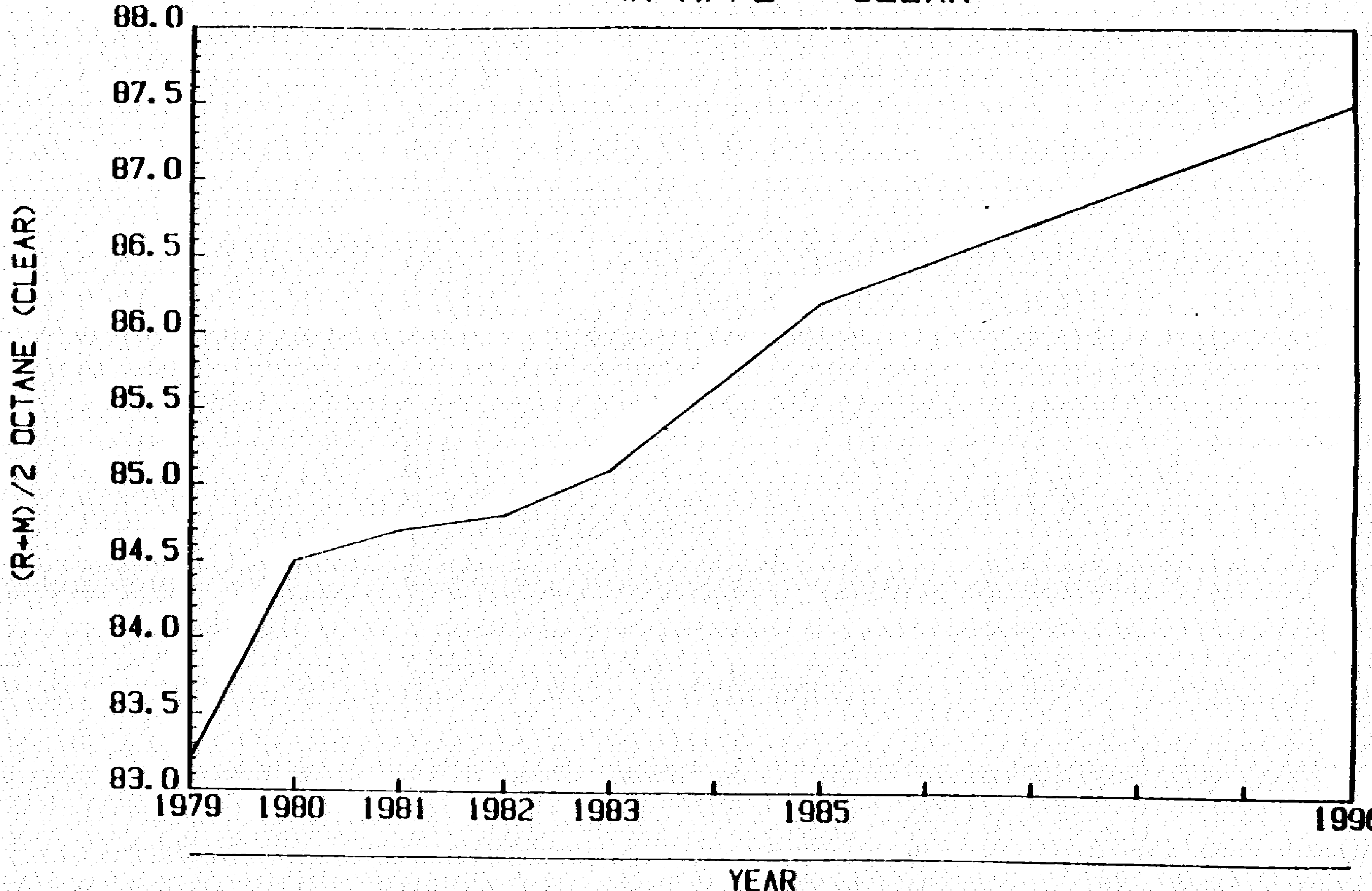


TABLE V-A-9

ESTIMATED U.S. DEMAND FOR TRANSPORT FUEL ETHANOL, 1981-1990
(Million Gallons)

1981	1982	1983	1985	1990
77	225	407	600	850

Chem Systems' forecast indicates that ethanol will represent approximately one-quarter of the total consumption of octane enhancers used by U.S. refiners in the late 1980s and 1990. Total enhancer use, including MTBE, TBA, methanol, dimate and ethanol, is projected to grow from less than 50 MBPCD in 1980 and about 100 MBPCD in 1983 to 160 MBPCD in 1985 and just over 225 MBPCD in 1990. The ethanol share of this market will grow from about 10 percent in 1980 to an estimated 25 percent by 1990 as shown in Table V-A-10. In order to achieve this market share additional ethanol facilities will be required.

TABLE V-A-10

OCTANE ENHANCER USAGE FOR GASOLINE BLENDING, 1980-1990
(Percent of Total)

	1980	1983	1985	1990
Ethanol	10	23	24	25
MTBE	15	18	15	15
Methanol and higher alcohols	40	24	30	33
Dimate	35	35	29	26
Total	100	100	100	100

Regional and State Gasoline Demand

Regional gasoline demand in Michigan and surrounding states is summarized in Table V-A-11. In recent years, the growth rate in Michigan has outpaced that of the region as a whole and the national average.

TABLE V-A-11

GASOLINE DEMAND BY STATE

	1982 (MBPCD)	Growth Rates, % 1960-1978	1978-1982
Michigan	237	3.5	(7.2)
Minnesota	95	2.2	(6.4)
Wisconsin	100	3.1	(8.9)
Indiana	150	2.7	(5.3)
Illinois	252	2.9	(6.6)
Ohio	282	2.8	(4.9)
National total	6,539	3.6	(4.4)

Furthermore, there is a growing regional demand for unleaded fuel. This is exemplified in Table V-A-12 which summarizes Chem Systems' demand projections for the state of Michigan. It can be seen that while the total gasoline demand is projected to decrease throughout the study period unleaded gasoline will be growing rapidly until 1990. From that point, total unleaded will also begin to decline, but the premium grade will continue its growth. It is interesting to note that virtually the entire gasoline market in Michigan is within 200 miles of Gladwin County, the proposed plant site.

4. Tax Incentives

It has been a combination of federal and state legislative initiatives that have provided the impetus for the development of a domestic fermentation ethanol industry. Several of the federal programs have been specifically directed to benefit ethanol production and usage, while others have had a broader intent but nonetheless have directly supported the development of the domestic industry. Direct programs include the 5 cent per gallon federal exemption (increasing to 6 cents on January 1, 1985) for using blends of 10 percent ethanol and 90 percent gasoline as well as various loan guarantees for plant construction and so forth. Indirect programs include the phasedown of lead usage in gasoline as part of the Clean Air Act and various industrial tax incentives created as part

TABLE V-A-12

MICHIGAN GASOLINE DEMAND BY GRADE

	Million Gallons			
	Unleaded		Leaded	
	Regular	Premium	Regular	Total
1983	1,793	225	1,664	3,681
1985	2,130	355	1,065	3,550
1990	2,203	518	518	3,240
1995	2,085	556	139	2,780
2000	1,823	608	0	2,430

	Grade Distribution (Percent)			
	Unleaded		Leaded	
	Regular	Premium	Regular	Total
1983	48.7	6.1	45.2	100.0
1985	60.0	10.0	30.0	100.0
1990	68.0	16.0	16.0	100.0
1995	75.0	20.0	5.0	100.0
2000	75.0	25.0	0.0	100.0

Market Distribution from Gladwin County	
Miles	Percent of Michigan Demand
0-50	9.6
50-100	38.6
100-200	50.2
200-300	1.6
	100.0

of President Reagan's tax reform initiatives. This section will review the federal programs that provide direct support, touch on those indirect issues that impact on ethanol use and plant construction, and discuss the state supports for fuel ethanol use.

Direct Federal Incentives

Perhaps the best known incentive, and certainly one of the most important, is the exemption from part of the federal excise tax for fuels containing ethanol produced from biomass sources - the law specifically excludes ethanol produced from petroleum and natural gas derivatives. Therefore, "synthetic" ethanol produced from ethylene does not qualify for the federal credit (nor would it qualify for most state credits). The original legislation exempted 10 percent ethanol blends from the full 4 cent per gallon federal excise tax on gasoline and diesel fuel - this exemption was originally scheduled to last until October 1984. As part of the Crude Oil Windfall Profits Tax Act of 1979, the exemption was extended until December 31, 1992. During the closing days of the 97th Congress, the Surface Transportation Assistance Act of 1982 was passed. This act increased the federal excise tax to 9 cents per gallon and raised the level of the exemption for 10 percent anhydrous alcohol blends by 1 cent per gallon to 5 cents per gallon and continued it in place until December 31, 1992 (the bill was effective on April 1, 1983). The increase in the exemption was the result of legislative compromise. The Senate wanted to provide ethanol with a full 9 cent per gallon exemption while many in the House wanted to continue the 4 cent per gallon exemption and terminate it in 1988. More recent efforts have resulted in an increase of the level of federal tax exemption once again, from 5 cents to 6 cents per gallon, effective on January 1, 1985 and continuing until 1992.

Coincident with the increase in the federal tax exemption, the blender tax credit was also increased. Effective April 1, 1983, and again applicable to December 31, 1992, biomass-derived alcohol used in or as a motor fuel was eligible for a credit of 50 cents per gallon for alcohol of at least 190 proof, and a 37.5 cent per gallon credit applies for alcohol from 150 to 190 proof. These benefits must be reduced by any gained from the overall federal excise tax exemption. This credit will increase to 60 cents on January 1, 1985.

In addition to the above credits, there is a full 9 cent per gallon tax exemption for neat alcohol transport fuels. This applies to any alcohol fuel containing at least 85 percent methanol, ethanol and any other alcohol derived from either biomass and coal and specifically excluding natural gas or petroleum-derived products.

Construction Assistance

To aid the construction of ethanol plants, both the Department of Energy and the Farmers Home Administration have been providing loan guarantees. With the budget cutting of the current administration, the level of funding available for guarantees has dropped considerably. Also, the uncertainties associated with independently financing these projects, given the upcoming end of the Energy Investment Tax Credit, higher feedstock costs, lower ethanol netbacks, and declining gasoline prices, have delayed many proposed projects.

The Department of Energy (DOE) has only limited funding available to support new plant construction, while the Synthetic Fuels Corporation (SFC) is specifically excluded from funding biomass ethanol plants. A DOE loan guarantee has been provided to New Energy of Indiana, whose plant construction is virtually completed. Conditional guarantees have also been provided to a number of potential large-scale ventures, although sources of both equity and debt financing must be arranged before a commitment can be obtained. In addition, companies must provide a detailed marketing plan and design and engineering details before a guarantee will be granted.

Having a conditional loan guarantee is no assurance that a project will in fact go ahead. Many details must be worked out, including securing the necessary equity and getting debt financing. Many project sponsors have also sought additional governmental arrangements (federal, state, and local) to secure their projects. Among the options are the following:

- Small Business Administration Loans
- Urban Development Action Grants (UDAG)
- Industrial Development Bonding

Energy Investment Tax Credit (EITC)

The tax codes now provide a 10 percent energy investment tax credit, which is in addition to the regular 10 percent investment tax credit for alternative energy projects such as biomass ethanol facilities. This credit is due to expire on December 31, 1985, and recent attempts to extend the credit have been unsuccessful. Under current law a plant would have to be completed prior to January 1, 1986, in order to qualify for the EITC. The impending end of the credit has caused financing delays for several projects such as the proposed Minnesota Alcohol Fuel project. To qualify for the EITC the primary source of energy for the project must not be either oil or natural gas or derived from them.

Furthermore, if a project is funded with Industrial Development Bonds (IDB), the EITC will not be available. Also, if IDBs are used, the basis for depreciating property is changed.

Indirect Programs

The following are the most significant other federal programs that impact on ethanol:

- Farm support programs such as the Payments in Kind (PIK), which affects corn and other ethanol plant feedstock costs.

- The Clean Air Act and its lead phasedown requirements, which increases the demand for octane enhancers.

- Tax incentives that allow for accelerated cost recovery in the depreciation schedule and other similar issues.

Under the proposed regulations the level of lead would be reduced to 0.5 grams per gallon of leaded gasoline over an interim period such as from 1985-1987. After that time the lead level would drop to 0.1 grams per gallon. During the fourth quarter of 1984 final rules should be developed. The analysis that has been performed for this study is based on existing law although it is recognized that such action by the EPA would significantly add to the demand for ethanol and other octane enhancers. Accordingly there should be reasonable upside potential to the base numbers presented in this report.

State Incentives

State tax incentives enable ethanol-gasoline blends to compete in the marketplace. During 1982, thirty-one (31) states provided an incentive that varied from a low of 1 cent per gallon to as much as 10 cents per gallon in New Mexico. These incentives in most cases apply to alcohol from any biomass source, but in several states there are restrictions that reflect feedstock type and source, ethanol production location, and reciprocal agreements between states. In this section a brief review has been developed of the various changes in state exemptions in the Midwest that have occurred during 1983 and the first half of 1984 and may occur thereafter. A summary of the existing state-by-state exemptions is also provided in Table V-A-13. It must be emphasized that there is no certainty that these tax laws will remain unchanged, given the exigencies of state budgets, political pressures, etc. However, as discussed in this section, more states have in fact revised their laws to provide greater incentives than have reduced them. The continuation of these exemptions will be based on a combination of revenue loss implications, political pressure and the level of industrial development and job creation that can be attributed to these exemptions. A brief description of the Michigan situation follows.

Michigan

The state has maintained its phasedown ethanol exemption for local ethanol production. However, the state lowered its tax break, effective April 1, 1983, from 3 cents to 2 cents per gallon for blends containing ethanol produced out of state and in states that do not match Michigan's 4 cent per gallon tax exemption. The exemption for out-of-state material dropped to 1 cent per gallon in 1984 and expires on December 31, 1984.

TABLE V-A-13

SUMMARY OF STATE TAX EXEMPTIONS*

(All Units are Cents per Gallon Except as Noted)

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	In-state Production Only	Reciprocity
Alabama	3	3	3	3	3	3	3	3	3	3	3	x	x
Alaska	8	8	8	8	8	8	8	8	8	8	8		
Arizona	-	-	-	-	-	-	-	-	-	-	-		
Arkansas	6.5	6.5	5.5	5.5	5.5	5.5	5.5	5.5	5.5	5.5/0	-	x	x
California	4	3	3/0	-	-	-	-	-	-	-	-		
Colorado	5	5	5	5/0	0	-	-	-	-	-	-	x	
Connecticut	1	1	1	1	1	1	1	1	1	1	1		
Delaware	-	-	-	-	-	-	-	-	-	-	-		
Florida	5	5%**/4	4	4	4	4/0	-	-	-	-	-		
Georgia	-	-	-	-	-	-	-	-	-	-	-		
Hawaii	4	4	4	4/0	-	-	-	-	-	-	-		
Idaho	4	4	4	4	4/0	-	-	-	-	-	-	x	
Illinois	3%	3%/2%/4%	4%	4%	4%	4%	4%	4%	4%	4%	4%	x	x
Indiana(1)	4%	3%	3%/2.5	2.5/0	-	-	-	-	-	-	-		
Iowa	5	5/3	3/2	2/1	1/0	-	-	-	-	-	-		
Kansas	2	2/3/4	5	5	5	5	5	5	5	5	5		
Kentucky	3.5	3.5	3.5	3.5	3.5/0	-	-	-	-	-	-	x	x
Louisiana	8	8	8/16	16	16	16	16	16	16	16	0	x	
Maine	-	-	-	-	4	4	4	4	0	0	0	x	
Maryland	-	0/3	3	3	3/0	-	-	-	-	-	-		
Massachusetts	-	-	-	-	-	-	-	-	-	-	-		
Michigan -													
local prod'n	5	4	4	2	1	-	-	-	-	-	-		
out of state	5	4/3/2	1	0	-	-	-	-	-	-	-		x
Minnesota	-	0/2	2	2/4	4	4	4	4	4	4	4/0		
Mississippi	-	-	-	0/6%	6%	6%	6%	6%	6%	6%	6%	x	
Missouri	-	-	-	-	-	-	-	-	-	-	-		
Montana	7	7	7	7/5	5/3	3	3	3/0	-	-	-	x	
Nebraska	5	5	5	5	5	5	5	5	5	5	5	x	
Nevada	1	1	1	1	1	1	1	1	1	1	1		
New Hampshire	5	5/0	-	-	-	-	-	-	-	-	-	x	
New Jersey	-	-	-	0/8	8	8	6	6	4	4	0	x	
New Mexico	10	10/11	11	11	11	11/8	8/5	5/3	3/2	2/0	0	x	

*As of August 1984.
**Refers to percentage of wholesale price of gasoline.
x Applies to state only.

(1) State now offers a special ethanol production subsidy. Effective 7/1/84 Indiana ethanol producers will receive 10 cents per gallon of production and after 7/1/85 subsidy will increase to 25 cents per gallon.

TABLE V-A-13
(Continued)

SUMMARY OF STATE TAX EXEMPTIONS*

(All Units are Cents per Gallon Except as Noted)

	1982	1983	1984	1985	1986	1987	1988	1989	1990	1991	1992	In-state Production Only	Reciprocity
New York	-	-	-	-	-	-	-	-	-	-	-		
North Carolina	2	2/1/5	5	5	5	5	5	5	5	5	5/0		
North Dakota	4	4	5	6	4	4	4	4	4	4	4/0		
Ohio	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5		
Oklahoma	6.5	-	-	-	-	-	-	-	-	-	-		
Oregon	-	-	-	-	-	-	-	-	-	-	-		
Pennsylvania	-	-	-	-	-	-	-	-	-	-	-		
Rhode Island	-	-	-	-	-	-	-	-	-	-	-		
South Carolina	7	0	-	-	-	-	-	-	-	-	-		
South Dakota	4	4	4	4/0	-	-	-	-	-	-	-		x
Tennessee	-	4	4	4	4	4	4	-	-	-	-		x
Texas -													
local prod'n	5	5	5	5	5	4	3	2	1	0	0		
out-of-state	5	5/1.2	1.2	1.2	1.2	1.0	0.7	0.5	0.25	0	0		x
Utah	5	5	5	5	5	5	5	5	5	5	5		
Vermont	-	-	-	-	-	-	-	-	-	-	-		
Virginia	8	8	8/6	6	6/4	4	4/2	2	2/0	0	0	x	
Washington	1.5	1.5/2.6	2.6/2.9	2.9	2.9	0	-	-	-	-	-		
West Virginia	-	-	-	-	-	-	-	-	-	-	-		
Wisconsin	-	-	-	-	-	-	-	-	-	-	-		
Wyoming	4	4	4	0	-	-	-	-	-	-	-	x	

*As of August 1984.
**Refers to percentage of wholesale price of gasoline.
x-Applies to state only.

There are also two mutually exclusive efforts involving the state exemption. One would abolish the credit entirely, while another would double the credit and apply it only to alcohol produced from at least 50 percent Michigan-grown feedstocks. It is unclear what final action may occur on either of these measures, although it is considered likely that Michigan will extend its tax credit if industrial development occurs as a result.

B. Demand and Pricing Analysis

1. Ethanol

Ethanol Demand in Michigan

Ethanol usage in Michign for gasoline blending was 56.6 million gallons in 1983 which made it the number one consuming state of fuel ethanol in the U.S. Meanwhile, Michigan ethanol production was only 0.3 million gallons in 1983 and will expand to 12.5 million gallons by 1987.

Chem Systems has examined the demand for gasoline in Michigan and has developed both volume projections as well as demand by grade. On the basis of these projections, ethanol demand assessments have been developed for various levels of market penetration and for varying distances from the plant site. These analyses were developed using existing lead use regulations. If an accelerated lead phasedown program is enacted, there will be even greater demand for ethanol. Overall gasoline volume demand projections by grade over the study period have been presented in Table V-A-12. Using these data as the basis the potential for ethanol was developed.

A profile of ethanol demand by year, radius from the Gladwin County plant site and penetration level is shown in Tables V-B-1 to V-B-5. In terms of unleaded premium, the proposed 25 million gallon per year ethanol plant would have to penetrate 48 percent of the potential 1990 market, 45 percent of the 1995 market and 41 percent of the 2000 market in Michigan.

However, in terms of total unleaded regular plus premium, the proposed plant would only have to penetrate 9 percent of the potential 1990 market and slightly more in future years. This should be an easy goal. In fact it should be reasonable for half the plant output to find its way into premium and the other half into regular.

TABLE V-B-1

ETHANOL MARKET POTENTIAL, 1983
(Million Gallons)

Distance from Gladwin site, miles	0-50	50-100	100-200	200-300	Total
Gasoline					
Unleaded regular	172	692	900	29	1,793
Unleaded premium	22	87	113	4	225
Leaded regular	160	642	835	27	1,664
Total					3,681
Potential Ethanol Penetration					
100% of ULR & ULP	19.4	77.9	101.3	3.2	201.7
100% of ULP	2.2	8.7	11.3	0.4	22.5
50% of ULR & ULP	9.7	38.9	50.6	1.6	100.9
25% of ULR & ULP	4.8	19.5	25.3	0.8	50.4

TABLE V-B-2

ETHANOL MARKET POTENTIAL, 1985
(Million Gallons)

Distance from Gladwin site, miles	0-50	50-100	100-200	200-300	Total
Gasoline					
Unleaded regular	204	822	1,069	34	2,130
Unleaded premium	34	137	178	6	355
Leaded regular	102	411	535	17	1,065
Total					3,550
Potential Ethanol Penetration					
100% of ULR & ULP	23.9	95.9	124.7	4.0	248.5
100% of ULP	3.4	13.7	17.8	0.6	35.5
50% of ULR & ULP	11.9	48.0	62.4	2.0	124.3
25% of ULR & ULP	6.0	24.0	31.2	1.0	62.1

TABLE V-B-3

ETHANOL MARKET POTENTIAL, 1990
(Million Gallons)

Distance from Gladwin site, miles	0-50	50-100	100-200	200-300	Total
Gasoline					
Unleaded regular	212	850	1,106	35	2,203
Unleaded premium	50	200	260	8	518
Leaded regular	50	200	260	8	518
Total					3,240
Potential Ethanol Penetration					
100% of ULR & ULP	26.1	105.1	136.6	4.4	272.2
100% of ULP	5.0	20.0	26.0	0.8	51.8
50% of ULR & ULP	13.1	52.5	68.3	2.2	136.1
25% of ULR & ULP	6.5	26.3	34.2	1.1	68.0

TABLE V-B-4

ETHANOL MARKET POTENTIAL, 1995
(Million Gallons)

Distance from Gladwin site, miles	0-50	50-100	100-200	200-300	Total
Gasoline					
Unleaded regular	200	850	1,047	33	2,085
Unleaded premium	53	215	279	9	556
Leaded regular	13	54	70	2	139
Total					2,780
Potential Ethanol Penetration					
100% of ULR & ULP	25.4	101.9	132.6	4.2	264.1
100% of ULP	5.3	21.5	27.9	0.9	55.6
50% of ULR & ULP	12.7	51.0	66.3	2.1	132.1
25% of ULR & ULP	6.3	25.5	33.1	1.1	66.0

TABLE V-B-5

ETHANOL MARKET POTENTIAL, 2000
(Million Gallons)

Distance from Gladwin site, miles	0-50	50-100	100-200	200-300	Total
Gasoline					
Unleaded regular	175	703	915	29	1,823
Unleaded premium	58	234	305	10	608
Leaded regular	0	0	0	0	0
Total					2,430
Potential Ethanol Penetration					
100% of ULR & ULP	23.3	93.8	122.0	3.9	243.0
100% of ULP	5.8	23.4	30.5	1.0	60.8
50% of ULR & ULP	11.7	46.9	61.0	1.9	121.5
25% of ULR & ULP	5.8	23.4	30.5	1.0	60.8

Ethanol Valuation

Ethanol has the highest value as an octane enhancer of the various alcohols and ethers now being utilized by refiners as well as gasoline blenders and marketers. Its use also exceeds that of any other single enhancer commonly utilized. However, because of its relatively high cost ($1.55-$1.70 per gallon), its use is largely a function of both federal and state tax credits granted for the use of 10 percent ethanol in gasoline blends.

The value of an octane enhancer to the refining industry is based on:

- The cost of crude oil
- The price of gasoline
- The value of butane
- The cost of incremental octane

For most octane enhancers, the traditional method of assessing their value is based on their physical properties compared to a gasoline reference. The properties of interest are octane level and vapor pressure. Both blending octane number, (R+M)/2, and the blending Reid vapor pressure (RVP) are significantly influenced by the gasoline base stock composition (normally defined by the PONA analysis) and lead level. Furthermore, with the exception of MTBE, the blending RVPs of the other normally used oxygenates exhibit severe deviation from expected vapor pressures as predicted by Raoult's Law. These deviations are all on the up side which equates to a butane backout factor in order to maintain a specified gasoline RVP when oxygenates are blended with gasoline.

Ethanol has a blending octane of about 119 compared to the reference gasoline value of 88. Because of this difference of 31 octane numbers, ethanol has a significant octane credit associated with it. Also, its blending vapor pressure (RVP) is 21.8 psi in a 10 percent blend compared to a gasoline reference value of 11.

This gasoline reference vapor pressure represents an approximate national average. There are both seasonal and geographic variations that are due to local climatic conditions. The higher RVP of ethanol means that refiners would have to back out butane from their gasoline to bring the vapor pressure up to specification. Since butane is valued at less than gasoline, the refiner effectively loses value. In this instance the enhancer would receive a butane debit to reflect this downgrading were it not for special regulations.

In effect there is no butane debit since ethanol, at up to 10 volume percent, has a specific environmental exemption to allow a higher vapor pressure of the finished gasoline blend. This exemption does not currently apply to any other enhancer. Under most current practice, therefore, ethanol is not blended at the refinery but rather at the jobber/distributor level. Ethanol is generally added to regular-unleaded-grade gasoline to produce a premium-grade unleaded gasoline with a higher vapor pressure.

Accordingly, ethanol has two intrinsic values:

- One based on blending at the terminal which only takes into consideration the octane credit.

- The other applies if the blending occurs at the refinery and butane is backed out. Refiners have the option of meeting the vapor pressure specification, in which case they would back out butane. They have no legal requirement to do so, but some wish to maintain constant quality specifications. A refiner could also choose an alternative approach, which involves the preparation of a base stock at the refinery with a lower vapor pressure than normal. This base stock would then be used for blending purposes at the terminal.

The difference in ethanol value for these situations is about 5 cents per gallon currently and is expected to rise to about 9 cents per gallon by 1990. In a Midwest location offering a combined tax credit of 9 cents per gallon of blend (5 cents federal and 4 cents state), ethanol would be worth about $1.69 per gallon in 1983 assuming that the ethanol is used in an unleaded-regular-grade gasoline. If unleaded regular is upgraded to premium gasoline by the addition of 10 percent ethanol, the ethanol value increases to approximately $2.12 per gallon.

Chem Systems has developed several ethanol valuation cases for this study. The key factors in determining the netback to the ethanol plant are freight costs and the level of state tax credit which has been assumed as 3 cents per gallon from 1985 onward.

In developing the netback values Chem Systems also performed the following:

- Typical Midwestern terminal contract unleaded regular and premium prices were developed. While it is recognized that there will be some slight variation in terminal prices between various markets, it is deemed appropriate to use a typical value for feasibility study purposes.

- There would be some discount required to give the jobber/blender/terminal operator an incentive to handle the additional product. The incentive used for calculation purposes was 1.1-2.0 cents per gallon of blend over the 1983-1995 period. If an ethanol producer also owned distribution outlets he would "capture" this incentive.

A typical calculation for a Michigan-only marketing case is illustrated for 1985 in Table V-B-6. Three cases were developed:

- Case I illustrates the value of ethanol in unleaded regular gasoline. While on a calculated basis, this represents the lowest value for ethanol, under current market conditions this tends to more closely reflect market clearing conditions.

- Case II reflects ethanol value when blended into unleaded premium gasoline.

- Case III represents the ethanol value when used to upgrade unleaded regular to unleaded premium gasoline. This represents the highest ethanol value, but market clearing transactions do not occur at this level. However, this value can be realized by an ethanol producer who also owns gasoline distribution outlets (either wholesale or retail).

A summary of the ethanol netback calculations for 1983-2000 appears in Table V-B-7. This table shows the three sets of values. The maximum value shown represents the ethanol netback value when used for upgrading regular unleaded gasoline to unleaded premium. The minimum case shows the ethanol value in unleaded regular gasoline and is more typical of market clearing transactions. The midrange case would reflect the ethanol value for boosting unleaded premium from 90 to 93 octane.

TABLE V-B-6

ETHANOL VALUATION CALCULATION, 1985
(Cents per Gallon)

I. Value of Ethanol in Unleaded Regular

Unleaded regular gasoline (at terminal)	90.7
Terminal incentive	1.2
Net blend value	89.5
Federal incentive	6.0
State incentive	3.0
Total incentives	9.0
Implied ethanol value at terminal = 10((89.5 - .9(90.7)) + 9.0) =	163.7
Average freight	3.6
Ethanol netback	165.1

II. Value of Ethanol in Unleaded Premium

Unleaded premium gasoline (at terminal)	96.5
Terminal incentive	1.2
Net blend value	95.3
Federal incentive	6.0
State incentive	3.0
Total incentives	9.0
Implied ethanol value at terminal = 10((95.3 - .9(96.5)) +9.0) =	174.9
Average freight	3.6
Ethanol netback	170.9

III. Value of Ethanol in Upgrading Unleaded Regular to Unleaded Premium

Unleaded regular gasoline (at terminal)	90.7
Unleaded premium gasoline (at terminal)	96.5
Terminal incentive	1.2
Net blend value	95.3
Federal incentive	6.0
State incentive	3.0
Total incentives	9.0
Implied ethanol value at terminal = 10((95.3 - .9(90.7)) +9.0) =	226.7
Average freight	3.6
Ethanol netback	223.1

TABLE V-B-7

ETHANOL VALUATION WITH MICHIGAN ONLY MARKETING
(Cents per Gallon)

	1983	1985	1986	1990	1995	2000
Unleaded Regular						
Midwest unleaded reg. MOGAS	90.3	90.7	92.0	118.3	156.0	202.0
Terminal incentive	1.1	1.2	1.3	1.5	2.0	3.0
Avg. ethanol transportation	3.3	3.6	3.7	4.8	6.5	8.4
Federal tax incentive	5.0	6.0	6.0	6.0	6.0	6.0
State tax incentive	4.0	3.0	3.0	3.0	3.0	3.0
Delv'd ethanol value	169.3	168.7	169.0	193.3	226.0	262.0
Ethanol netback (minimum)	166.0	165.1	165.3	188.5	219.5	253.6
Unleaded Premium						
Midwest unleaded prem. MOGAS	94.6	96.5	97.8	124.9	163.0	210.0
Delv'd ethanol value	173.6	174.5	174.8	199.9	233.0	270.0
Ethanol netback (midrange)	170.3	170.9	171.1	195.1	226.5	261.6
Ethanol upgrade value	212.3	226.7	227.0	259.3	296.0	342.0
Ethanol netback (maximum)	209.0	223.1	223.3	254.5	289.5	333.6

For this feasibility study, it has been assumed that half of the ethanol product would go into the upgrading of unleaded regular to unleaded premium and half would go into the unleaded regular market. Thus, as summarized in Table V-B-8, the ethanol netback values at the plant gate are the average of the minimum and maximum cases previously described.

TABLE V-B-8

ETHANOL NETBACKS, 1983-2001
(Cents per Gallon)

	Minimum	Midrange	Maximum	Plant Average
1983	166.0	170.3	209.0	187.5
1985	165.1	170.9	223.1	194.1
1986	165.3	171.1	223.3	194.3
1987	170.8	176.8	230.8	200.8
1990	188.5	195.1	254.5	221.5
1995	219.5	226.5	289.5	254.5
2000	253.6	261.6	333.6	293.6
2001	261.0	269.2	343.2	302.1

2. Furfural

Current Furfural Demand

The principal use for furfural is as an intermediate for the manufacture of furfuryl alcohol (FA), and to a lesser degree, tetrahydrofuran (THF). Furfural is also used as a selective solvent for the separation of saturated from unsaturated compounds in petroleum lubricating oil, and in the extractive distillation of butadiene.

In 1983, total U.S. consumption of furfural was 71 million pounds as shown in Table V-B-9. The major applications for furfural and its derivatives are discussed in the paragraphs which follow.

TABLE V-B-9

U.S. FURFURAL CONSUMPTION, 1983
(Million Pounds)

Markets		Consumption
Furfural derivatives		51
Furfuryl alcohol-foundry resins	28	
Tetrahydrofuran (THF)	23	
Lube oil refining		10.2
Butadiene extraction		1.8
Miscellaneous uses		8
Total		71

Furfuryl Alcohol

Furfuryl alcohol is used in resin form as a sand binder by the foundry industry, chemical resistant grouts and cements, and sand consolidation in oil well drilling. It is a mobile liquid which polymerizes readily under acid conditions. In addition to reacting with itself, FA is commercially co-reacted with aldehydes (formaldehyde), urea, and phenol.

The major market for FA resins is their use as binders for the production of molds and cores used for casting metals. Such resins were introduced as binders in the late 1950s and are used in two different metal casting systems - no-bake and hot-box.

In 1974, the consumption of FA resins amounted to about 70 million pounds with 30 million pounds being used in hot-box systems and 50 million pounds in no-bake applications. Based on an average furfuryl alcohol content of 75 percent for no-bake resins and 35 percent for hot-box resins, the consumption of furfuryl alcohol for foundry resins amounted to 48 million pounds in that year.

Although the prospects for growth looked good, the market for FA resins in the foundry industry actually decreased rather than increased in the period since 1974. By 1983, only 36 million pounds of FA resin were used by the foundry industry. Of this amount, 33 million pounds were used in no-bake systems and 3 million pounds in hot-box systems.

The reasons for this decline are as follows:

- The 6.7 percent per year decline in the shipments of foundry castings which occurred in the 1974-1983 period.

- The shortage of furfuryl alcohol that developed in the mid-1970s.

- Several price increases for furfuryl alcohol which effectively doubled the price of the material between 1974 and 1983. This undoubtedly had a dampening effect on the use of FA resin because the price of phenol, the key raw material in competing resins, only increased about 20 percent over the same period.

- The declining importance of hot-box systems as foundries switched from hot-box to cold-cure systems. This change was generally attributed to the energy intensive nature of the former rather than any advantageous technical advantage of the latter.

The future outlook for FA resins will depend on the growth of the foundry industry and the international competition between FA resins and other no-bake systems. Most industry representatives indicated that there is little opportunity for any recovery in the market for hot-box systems and, even if there were, FA resins have little change of regaining the market they lost to phenolic resins. Chem Systems estimates that FA resin consumption should increase to 50 million pounds by 1990 with no-bake systems using 47 million pounds and hot-box systems 3 million pounds. These projections are sensitive to price changes for the furfuryl alcohol and the effect of such changes are discussed later.

The demand for FA resins and furfuryl alcohol for 1983-2001 is presented in Table V-B-10.

TABLE V-B-10

DEMAND FOR FA RESINS IN FOUNDRY CASTINGS
(Million Pounds)

	1983	1984	1985	1990	1995	2001
FA resins						
No-bake	36	40	43	47	51	56
Hot-box	3	3	3	3	3	3
	39	43	46	50	54	59
Furfuryl alcohol	28	31	33	36	40	44
Furfural	28	31	33	36	40	44

The major producers of foundry resins are Ashland Chemical, Acme Resin, Delta Resins, Kordell and CE Cast Products. Ashland is the largest consumer of furfuryl alcohol and will probably use 7-8 million pounds this year. Delta is next in importance and its use of furfuryl alcohol will probably amount to 5-6 million pounds in 1984. Table V-B-11 lists the producers of foundry resins and their plant locations.

TABLE V-B-11

U.S. PRODUCERS OF FOUNDRY RESINS

Company	Plant Location
Acme Resin (CPC International)	Forest Park, Ill.
Ashland Chemical	Cleveland, Ohio
	Hammond, Ind.
Borden Chemical	Louisville, Ky.
CE Cast Industrial Products	Muse, Pa.
Core-Lube	Danville, Ill.
Delta	Milwaukee, Wis.
	Detroit, Mich.
Eronel	Hawthorne, Ca.
Georgia Pacific	Newark, Ohio
Mar-Cam (Hill & Griffith)	Hickory, N.C.
Durex (Hooker Chemical)	North Tonawanda, N.Y.
Kordell Industries	Mishawaka, Ind.
Reichhold Chemical	
Thiem	Milwaukee, Wis.
United Erie	Erie, Pa.

Tetrahydrofuran

The major derivative for tetrahydrofuran is polytetramethylene ether glycol (PTMEG), make commercially by DuPont and QO Chemicals. PTMEG is used in the production of a variety of thermoplastic and thermoset polymer systems. The most important are as follows:

- Thermoplastic polyurethane elastomers.
- Curable prepolymers for production of cast (thermoset) polyurethane elastomers.
- Spandex fibers.
- Copolyester-ether thermoplastic elastomers.

The THF demand for the 1983-2001 period is presented in Table V-B-12.

TABLE V-B-12

U.S. DEMAND FOR TETRAHYDROFURAN
(Million Pounds)

	1983	1985	1990	1995	2001
PTMEG	60	57	70	90	115
Solvent	31	34	38	40	42
Other	1	1	2	2	3
	92	92	110	132	160
Exports	15	19	20	25	30
Imports	-	-	-	-	-
	107	111	130	157	190
Furfural	23	23	23	28	35

The current producers of THF are presented in Table V-B-13. Three of the four producers of THF use 1,4-butanediol. The fourth, QO Chemicals makes THF from furfural at its plant in Memphis. QO Chemicals also converts a portion of its THF output into PTMEG.

TABLE V-B-13

U.S. PRODUCERS OF TETRAHYDROFURAN
(Million Pounds per Year)

Company	Location	Capacity
DuPont	LaPorte, Tex.	120
BASF Wyandotte	Geismar, La.	20
GAF	Linden, N.J.	15
QO Chemicals*	Memphis, Tenn.	15
		170

*THF derived from furfural.

Solvent Uses

One of the oldest applications for furfural is as a selective solvent for the refining of lube oil to increase its stability and improve viscosity index. The furfural refining process was developed by Texaco and is second in importance in the United States to phenol extraction. Furfural preferentially dissolves aromatic, unsaturated, sulfur and nitrogen compounds, thus reducing the oxidation sensitivity and changing the temperature/viscosity curve of the lube oil.

The quantity of furfural consumed by a refinery consists of filling the system initially, and the makeup to replace losses during recycling. In 1983, 10 million pounds of furfural were consumed in the United States for lube oil refining. The U.S. refiners with furfural extraction units are listed in Table V-B-14.

TABLE V-B-14

LUBE OIL REFINNING FACILITIES WITH FURFURAL EXTRACTION UNITS
(Barrels per Day)

Refinery	Location	Extraction Capacity
Amoco Oil	Casper, Wyoming	2,300
Ashland Oil	Catlettsburg, Kentucky	15,000
Cit-Con Oil	Lake Charles, Louisiana	28,000*
Conoco	Ponca City, Oklahoma	4,300*
Farmland Industries	Coffeyville, Kansas	3,800
Mobil Oil	Beaumont, Texas	18,300*
	Paulsboro, New Jersey	15,500
Quaker State Oil	Farmers Valley, Pennsylvania	1,800
	St. Marys, West Virginia	1,000
Shell Oil	Martinez, California	6,800**
Sohio	Lima, Ohio	6,000
Sun	Tulsa, Oklahoma	16,000
	Yabucoa, Puerto Rico	15,000
Texaco	Port Arthur, Texas	23,000**
Witco Chemical	Oildale, California	4,000
		160,300

*Includes Duo-Sol extraction capacity.
**Includes SO_2 extraction capacity.

The long-term prospects for furfural in lube oil refining are very modest, reflecting the maturity of this market. Chem Systems estimates that furfural consumption will probably increase 1-2 percent annually in the 1980s and 1990s. This is slightly lower than the projection for lube oil production in the 1980s because it anticipates the closing of 1-2 furfural extraction units in the 1985-1990 period and generally more efficient plant operation.

The furfural forecast for lube oil refining for 1983-2001 is presented in Table V-B-15.

TABLE V-B-15

U.S. FURFURAL CONSUMPTION IN LUBE OIL PRODUCTION
(Million Pounds)

	1983	1984	1985	1990	1995	2001
Furfural	10.2	10.6	10.8	11.5	12.5	13.0

Furfural is also used for the extractive distillation of butadiene to effect separation from mixed C_4 streams. This process was developed by Phillips Petroleum Company. In addition to furfural, other solvents are used commercially for butadiene extraction. Shell developed a process based on acetonitrile, while Exxon developed one based on cuprous ammonium acetate. in more recent years, newer processes based on other solvents have been developed that offer advantages of lower capital cost and utility requirements compared to the furfural and other systems.

The current consumption of furfural for butadiene extraction is estimated at 1.8 million pounds. Considering the long-term outlook for butadiene consumption in the United States, Chem Systems estimates that furfural used in solvent extraction will probably not exceed 2.0 million pounds by 1990. There is also the possibility that the existing extraction units could be switched from furfural to other solvents.

The demand for furfural in butadiene extraction is presented in Table V-B-16.

TABLE V-B-16

U.S. FURFURAL CONSUMPTION WITH BUTADIENE EXTRACTION
(Million Pounds)

	1983	1984	1985	1990	1995	2001
Furfural	1.8	1.8	1.8	2.0	2.0	2.0

Miscellaneous Uses

The consumption of furfural in miscellaneous uses amounted to about 10 million pounds in 1983. In some of the outlets the application calls for the use of furfural itself, while in others FA is used. Some uses are complicated by the fact that both FA and furfural are used depending upon the specific formulation. In certain instances, the product is used as a solvent, in others as a wetting agent/adhesive aid, and in still others the final product is a resin.

Future Furfural Demand

The markets for furfural and its derivatives consist of a number of mature applications for which no growth or relatively modest growth is projected for the study period. The demand for furfural is forecast to reach 83 million pounds in 1990 and 105 million pounds in 2001, representing an average annual growth of 2.0 percent. The consumption of furfural by end use for the 1983-2001 period is presented in Table V-B-17.

TABLE V-B-17

U.S. FURFURAL CONSUMPTION, 1983-2001
(Million Pounds)

Markets	1983	1984	1985	1990	1995	2001	Average Annual Growth, % 1983-2001
Furfural derivatives							
Furfural alcohol-foundry resins	28	31	33	36	40	44	2.5
Tetrahydrofuran	23	23	23	23	28	35	2.4
Lube oil refining	10.2	10.5	11	11.5	12.5	13	1.5
Butadiene extraction	1.8	2	2	2	2	2	0.0
Miscellaneous uses	8	9	10	10.5	10.5	11	0.5
	71	75.5	79	83	93	105	2.0

Furfural Supply

QO Chemicals is the only producer of furfural in the United States. QO Chemicals is a privately-owned company that purchased the Quaker Oats Chemical Division from Quaker Oats during the first half of 1984. QO Chemicals is essentially being operated by the former management of Quaker Oats Chemical.

The current capacity for furfural is 132 million pounds per year. QO Chemicals is currently operating plants at Omaha, Nebraska; Cedar Rapids, Iowa; and Belle Glade, Florida. QO Chemicals was operating a fourth plant at Memphis, Tennessee, but this facility was shutdown this summer. Its capacity was 40 million pounds per year. All of these plants have been in operation for a number of years, although each has probably undergone considerable maintenance. This is because of the corrosion problems inherent with the conversion of agricultural wastes to furfural.

In addition to the above, Quaker Oats operated a fifth plant at Bayport, Texas between 1976 and 1983. This facility had a capacity of 44 million pounds per year and used rice hulls as raw material. According to several sources, the plant never operated at capacity and was plagued by severe corrosion problems. This facility was closed prior to the sale of the division to QO Chemicals.

QO Chemicals is also the only domestic producer of furfuryl alcohol. These facilities are located at Omaha, Nebraska and Memphis, Tennessee. Quaker Oats also had a furfuryl alcohol plant at its Bayport plant site. Chem Systems believes this facility has been moved to one of the other two sites to replace one or more older units. QO Chemicals' furfuryl alcohol capacity is estimated to be in excess of 50 million pounds per year.

QO Chemicals' U.S. plants and their capacities are listed in Table V-B-18.

TABLE V-B-18

QO CHEMICALS U.S. FURFURAL PLANTS
(Million Pounds per year)

Location	Capacity	Agricultural Feedstock
Omaha, Nebraska	40	Corn cobs
Cedar Rapids, Iowa	20	Corn cobs, oat hulls
Belle Glade, Florida	72	Sugar cane bagasse
Total	132	

Due to the fact that the sources of supply of furfural are dominated by countries with developed agricultural economies, there is considerable international trade between these sources and the industrialized nations which are the primary consumers of furfural and its derivatives. The United States stands out as the major supplier and user of furfural.

Even though there has been very little publicity given to the South Puerto Rico Sugar Company, a subsidiary of Gulf & Western Industries, its plant in the Dominican Republic is a major source for furfural. The company does not convert furfural to FA or other derivatives.

Table V-B-19 shows that furfural imports increased substantially in 1977-1979, then declined to the 5-10 million pound level in 1980-1983. However, imports during the first seven months of 1984 have already exceeded that level, amounting to 11.5 million pounds. For the year, imports could amount to as much as 18-20 million pounds with no change in the current rate of shipments.

TABLE V-B-19

U.S. IMPORTS OF FURFURAL, 1976-1983
(Thousand Pounds)

Year	Dominican Republic	South Africa	Other	Total
1976	552	-	526	1,078
1977	8,615	-	3,504	12,119
1978	16,668	878	5,441	22,987
1979	24,230	3,121	1,587	28,938
1980	2,899	416	1,722	5,037
1981	3,614	484	251	4,349
1982	5,489	2,141	38	7,668
1983	6,363	2,134	682	9,179

The importance of South Puerto Rico Sugar as a world supplier of furfural cannot be overlooked. Their nameplate capacity is 72 million pounds per year, which places them second to QO Chemicals as a free world producer. However, South Puerto Rico Sugar has been extremely reluctant to discuss any aspect of its furfural business. Since its basic business is sugar, it is highly likely that all of the furfural the company produces is sold through an intermediary.

The other important overseas source of furfural is SmithChem in South Africa. This company has been exporting furfural to the United States since 1978 with shipments in the last two years amounting to just over 2 million pounds. Several industry contacts noted that SmithChem has recently expanded its furfural plant and intends to increase its market position in the United States. However, this will more than likely result from an increase in furfuryl alcohol rather than furfural sales.

Data on U.S. exports of furfural are not available. However, the destination of U.S. exports in past years include Western Europe, Japan, Canada and Latin America. Chem Systems believes that U.S. exports may have amounted to 30-35 million pounds in the mid-1970s and 10-15 million pounds in the late 1970s. However, considering the global overcapacity in furfural and the increased level of competition in world markets, it is believed that U.S. exports are probably in the range of 5 million pounds in the last couple of years.

Furfural Pricing

Furfural is quite readily available in the United States and should remain so for all of the forecast period. QO Chemicals will probably not operate much above 50 percent of effective capacity in 1984 and, given the availability of furfural worldwide, can not be expected to improve this level of production in the next several years as shown in Table V-B-20.

TABLE V-B-20

U.S. FURFURAL SUPPLY/DEMAND BALANCE
(Million Pounds)

	1983	1984	1985	1990	1995	2001
Nameplate capacity	172	152	132	132	132	132
Effective capacity*	154	134	114	114	114	114
Domestic demand	71	75.5	79	83	93	105
Exports	5	5	5	-	-	-
Imports	9	20	10	10	12	14

*Based on Belle Galde plant operating 9 months a year.

This potential surplus supply of furfural in the United States should continue to exist unless one or more of the following changes occurs in the U.S. market.

- QO Chemicals closes one or more of the three furfural plants it is now operating.

- The market for furfural undergoes a significant change resulting in increased demand for either furfural or its derivatives.

Of the two, the former is likely to occur if no real change takes place in the outlook for furfural, or by-product furfural becomes available as a result of the cellulose-based manufacture of ethanol. The later is likely to occur if the market price for furfural is significantly reduced, opening up opportunities in existing or developing applications for furfural.

Two distinct opportunities exist to increase the demand for furfural beyond that forecast in Table V-B-20. Both are directly related to the sensitivity of the markets for foundry resins and THF to changes in the price of furfural.

The two key raw materials in the hot-box and no-bake resin segment of the foundry resin market are furfuryl alcohol and phenol. Currently, the market price for furfuryl alcohol is 62-64 cents a pound to large volume users compared to 30-31 cents a pound for phenol. This difference is directly related to the difference in the price of FA and phenolic resins. FA resins are now selling for about 75 cents a pound while phenolic resins sell for 30-40 cents a pound. Although some foundry resin producers are selling modified FA resins for 50-55 cents a pound in an attempt to offset this disadvantage, the modified resins lack the performance characteristics desired by most foundry operators.

All of the foundry resin producers agree that the foundry resin market is sensitive to raw material price changes. The most competitive segment of the market are the phenolic no-bake and the FA no-bake resins. Here, changes in raw material prices could result in significant changes in future demand. Although FA hot-box resins have lost most of this market to phenolic hot-box resins, this is a segment of the market where pricing could also bring about changes.

Chem Systems believes that FA resin demand in 1990 could increase by 30 million pounds compared to 1984 demand if the price of furfuryl alcohol would be reduced to 45 cents a pound (1984 constant dollars). The demand could probably be increased by a further 15 million pounds if the price of furfuryl alcohol could be reduced to 40 cents a pound (1984 constant dollars). Such prices would be equivalent to a furfural price of 30 cents a pound in the former case and 25 cents a pound in the latter case, both expressed as 1984 constant dollars.

Table V-B-21 presents the demand for FA resins based on furfuryl alcohol being sold at 45 cents a pound (constant 1984 dollars). The revised demand forecast is based on the following assumptions:

- By-product furfural would not become available until 1987-1988. As a result, the impact of lower furfuryl alcohol prices would not be reflected in the forecast until 1990.

- The market adjustment to lower furfuryl alcohol prices would be completed by 1990 and the demand for FA resins in subsequent years would reflect the maturity of the foundry resin market with no further shift in demand patterns.

- The primary penetration of FA resins would be felt by phenolic no-bake and hot-box resins. However, FA resins would also replace about 10 percent of the phenolic urethane cold-box and no-bake resins.

TABLE V-B-21

DEMAND FOR FA RESINS IN FOUNDRY CASTING - REVISED
(Million Pounds)

	1983	1984	1985	1990	1995	2001
FA resins						
No-bake	36	40	43	63	68	74
Hot-box	3	3	3	18	19	21
Other	-	-	-	3	4	4
	39	43	46	84	91	99
Furfural alcohol	28	31	33	56	61	66
Furfural	28	31	33	56	61	66

A reduction in the price of furfural should also put furfural-based THF on an equivalent cash cost basis with that based on 1,4-butanediol. This should enable QO Chemicals to compete more effectively in the domestic and export segments of the THF market. If QO Chemicals could double its market share in 1990, the market for furfural would increase by 23 million pounds.

In summary, the net effect of reducing the price of furfural in 1990 to 30 cents a pound (1984 constant dollars) would be to increase the domestic demand from 83 million to 126 million pounds in 1990.

The current and future pricing of furfural is summarized in Table V-B-22, showing prices for both market situations, i.e., with and without the availability of by-product furfural from wood-based ethanol facilities.

TABLE V-B-22

U.S. FURFURAL PRICES
(Cents Per Pound)

	1984	1985	1990	1995	2001
Without by-product Furfural available	47	47	55	61	67
With by-product Furfural available	-	-	40	51	58.5

3. Carbon Dioxide

There are many potential uses for the carbon dioxide by-product from a biomass-based ethanol plant. Figure V-B-1 shows historical carbon dioxide demand and how it is likely to change in the late 1980s.

Growth in demand for traditional carbon dioxide end uses will be substantial. The largest end use historically has been in urea production, although carbon dioxide for this use is usually supplied captively from integrated ammonia plants.

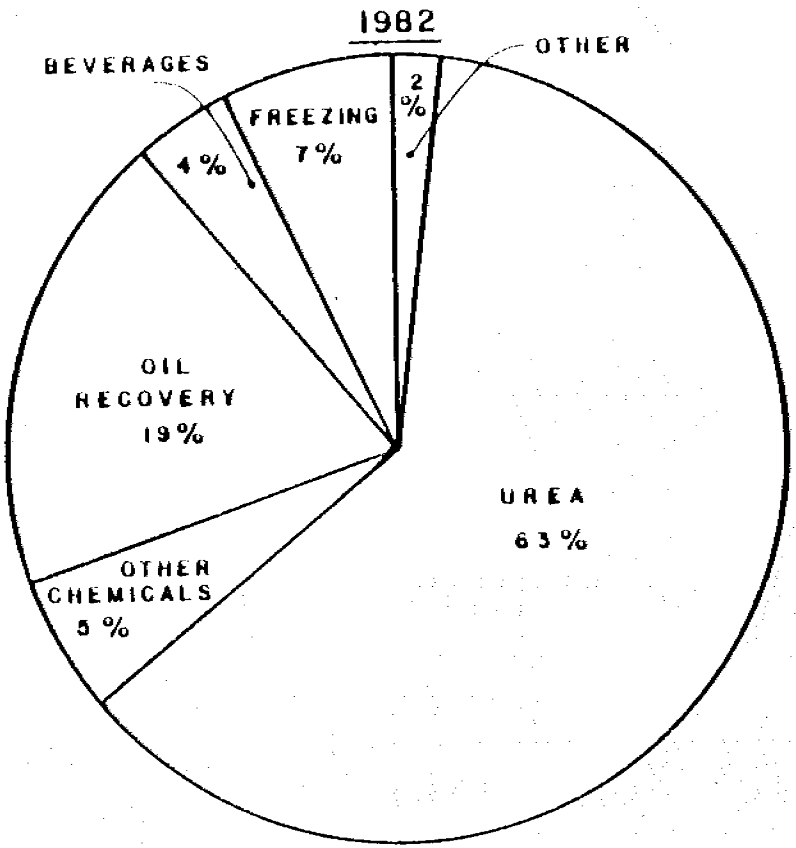
Carbon dioxide is used in food freezing where it competes with nitrogen for markets. This application has grown sharply over the last decade and is expected to continue to do so, especially to supply frozen foods for the fast-food industry and for home microwave cooking. Consumption of carbon dioxide for beverage carbonation has also increased rapidly in the past. However, as a maturing market, demand will probably rise to close to levels for other consumer-related goods.

Other areas of demand in the chemical process industry include use as an inerting agent and as a raw material in the production of methanol, and sodium and other inorganic salt carbonates. In methanol production from natural gas, carbon dioxide is reacted with excess hydrogen to make up for carbon deficiencies in the synthesis reaction feed.

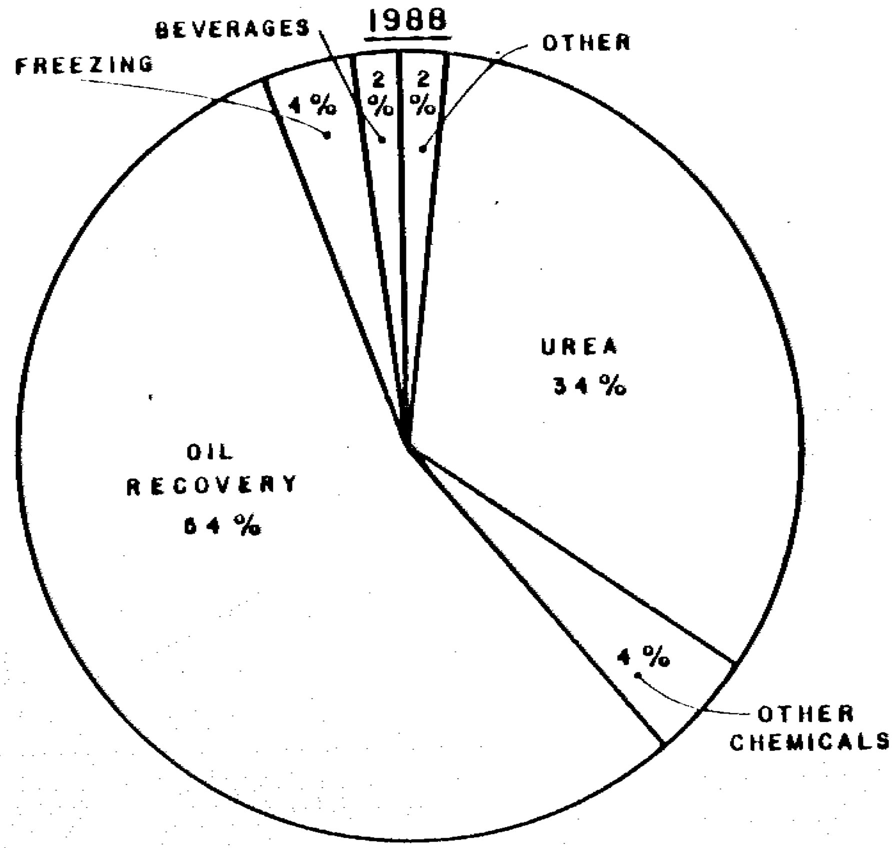
The predominant source of domestic carbon dioxide (Figure V-B-2) has historically been ammonia plants, which yield extremely high purity carbon dioxide. But high energy and feedstock costs, high interest rates and bumper grain crops have severely impacted the U.S. fertilizer and ammonia industry. Within the past three years, approximately 20 percent of domestic ammonia capacity has been shut down, which has had a major effect on carbon dioxide supplies.

FIGURE V-B-1

CO_2 DEMAND*



TOTAL = 46.4 THOUSAND SHORT TONS/DAY

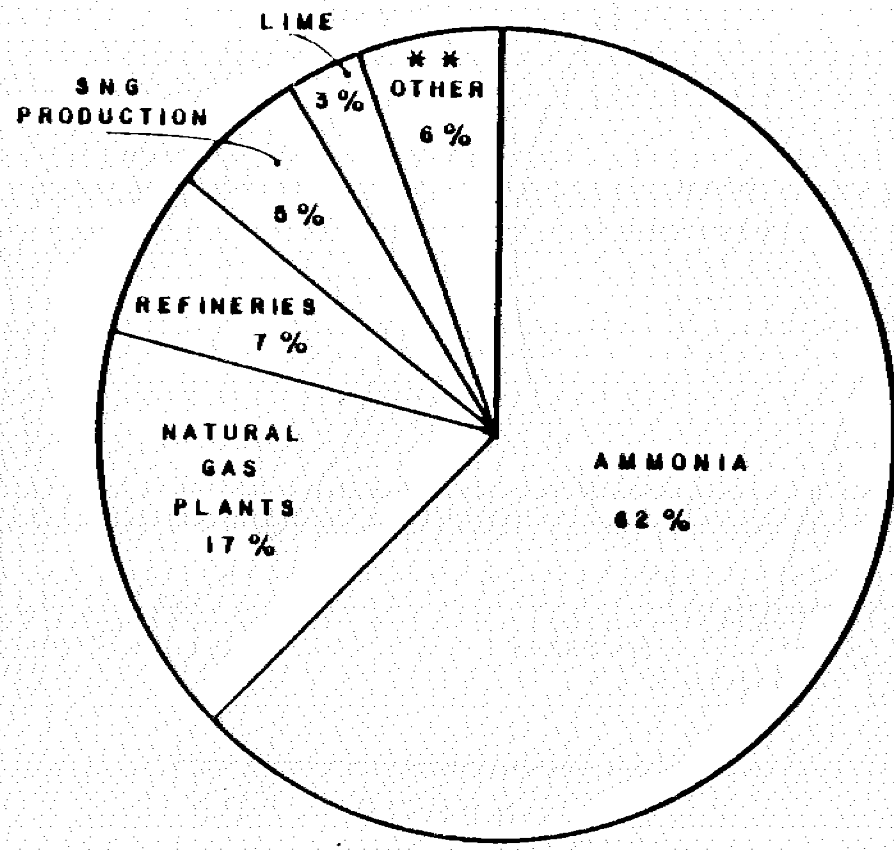


124.8 THOUSAND SHORT TONS/DAY

* PRELIMINARY ESTIMATES

FIGURE V-B-2

1982 CO_2 RECOVERY CAPACITY*



1982 TOTAL = 53.6 THOUSAND SHORT TONS / DAY

* PRELIMINARY ESTIMATE

** NATURAL DEPOSITS, ETHYLENE OXIDE, FERMENTATION, COMBUSTION FLUE GASES, SODIUM PHOSPHATE

By the end of this decade, the largest single source of carbon dioxide will be high-purity natural deposits in the Rocky Mountains, which will be developed to supply new EOR projects in west Texas and New Mexico. This material will be delivered via pipelines now under construction or being planned. As these deposits are tapped, they may become important sources of carbon dioxide for alternate markets, as well. Furthermore, as the EOR projects are operated, they will produce perhaps one-quarter to one-half of the injected carbon dioxide in associated production. This CO_2 can be reinjected in the same or nearby projects which will come on stream later.

Carbon dioxide off-gas from a fermentation ethanol plant can be a highly desirable product, depending on location. Many industrial gas companies such as Airco, Cardox, Liquid Carbonics, UGI actively seek out such sources of carbon dioxide and are prepared to enter into a long-term supply contract. These companies will typically install a liquefaction plant adjacent to the ethanol plant to minimize compression costs and then market the liquid carbon dioxide product usually for beverage carbonation.

For a Midwest location, the industrial gas companies have been paying 5-7 dollars per ton for the raw carbon dioxide stream. These are typically ten- to fifteen-year supply contracts with a built-in escalation tied in some manner to the ultimate price received for the liquid carbon dioxide product. A carbon dioxide liquefaction plant in the 300 ton per day capacity range will require an investment of about 6 million dollars. The cost of production, including a return on investment, will be in the 50-60 dollar per ton range, with the raw carbon dioxide gas priced in at 5-7 dollars per ton. Table V-B-23 summarizes the manufacturing cost for carbon dioxide liquefaction on a mid-1984 basis. Table V-B-24 shows a price forecast for raw carbon dioxide by-product assuming an arrangement is made with one of the companies mentioned above and wherein the price escalates with general inflation. Also shown in Table V-B-24 is a price forecast for liquefied carbon dioxide as projected for the Michigan ethanol facility.

TABLE V-B-23

LIQUEFIED CARBON DIOXIDE MANUFACTURING COST SUMMARY
(300 Tons per Day, mid-1984)

	Cents per Pound
Raw materials (1)	0.36
Utilities	0.42
Operating cost	0.38
Overhead expenses	0.39
Cash cost of production	1.56
Depreciation	0.65
Net cost of production	2.22
Carbon dioxide sales price at 10 percent DCF return	2.8

(1) Raw carbon dioxide at $6 per ton.

TABLE V-B-24

CARBON DIOXIDE BY-PRODUCT PRICE FORECAST
(Dollars per Ton)

Year	Raw CO_2 Gas	Liquefied CO_2
1987	6.30	64.40
1988	6.60	67.60
1989	6.90	70.60
1990	7.30	74.80
1991	7.60	77.80
1992	7.95	81.40
1993	8.30	85.00
1994	8.70	89.20
1995	9.10	93.20
1996	9.50	97.40
1997	9.90	101.40
1998	10.35	106.00
1999	10.80	110.60
2000	11.30	115.80
2001	11.80	120.80

VI. ECONOMIC ANALYSIS

A. Base Case Economics

1. Capital Cost Estimate

A capital cost has been estimated for a plant producing 25 million gallons per year of ethanol. The total inside battery limits (ISBL) cost is estimated at 66.26 million dollars based on mid-1984 costs for a Michigan site location. The cost estimate is based on Icarus cost data along with vendor equipment costs for major items.

A detailed equipment list for all major equipment items is presented in Appendix B. This list includes the design specifications for each item and its estimated purchase price on a mid-1984 basis. A breakdown of total ISBL capital cost by section is given in Table VI-A-1. Section 700, Carbon Dioxide Recovery System, is included in the ISBL as a complete turnkey facility.

The offsites include capital for storage, a steam/power cogeneration system, a wood handling facility, a cooling water system, waste treatment and other miscellaneous systems as described in Section IV-A-12. The offsites costs are summarized in Table VI-A-2.

2. Cost of Production Analysis

Table VI-A-3 is a production cost estimate for the base case 25 million gallon per year ethanol from wood plant. This estimate serves as the basis for the cash flow analysis presented in the following section. The costs are based on a mid-1984 time frame for the Michigan site. Most of the items in the cost of production estimate are self-explanatory. The by-product quantities, raw material and utility consumptions per gallon of ethanol produced have been discussed in Section IV-D.

TABLE VI-A-1

ISBL* CAPITAL COST SUMMARY FOR BASE CASE, MID-1984

Section	Name	Purchased Equipment, $	Total Field Installed Cost, $
100	Pretreatment/Prehyrolysis	1,683,000	3,104,000
200	Sugar Separation and Neutralization	2,052,400	3,901,000
300	Enzyme Production	2,972,000	6,376,000
400	Enzyme Hydrolysis	4,245,000	7,069,000
500	Sugar Concentration	3,731,000	5,941,000
600	Fermentation	1,714,000	4,209,000
800	Ethanol Purification	1,153,000	4,111,000
900	Furfural Production	1,638,000	5,166,000
1000	Heat Generation	2,100,000	3,311,000
Total Purchased Equipment		21,288,400	
Total Field Cost			43,188,000
Engineering			4,180,000
Construction Overhead			4,850,000
Bare Plant Cost			52,218,000
Contingency			5,222,000
Contractors Fee			1,509,000
Special Charges			1,550,000
Total Facility Cost			60,490,000
CO_2 Recovery System			5,800,000
Total ISBL			66,290,000

*Section 1100, Waste Treatment, is included in offsites costs.

TABLE VI-A-2

OFFSITES CAPITAL COSTS

	Total Installed Cost, $ Mid-1984
Storage	
Ethanol (14 days, 1,020,000 gals)	676,800
Calcium hydroxide (14 days, 160,000 gals)	560,000
Sulfuric acid (14 days, 85,000 gals)	195,000
Caustic (14 days, 40,500 gals)	82,600
Corn steep liquor (14 days, 40,500 gal)	82,600
Furfural (14 days, 160,000 gal)	171,000
Sodium carbonate (14 days, 23,500 gal)	160,000
Total storage	1,928,000
Steam boiler (320,000 lb/hr, 1,200 psig steam, wood-fired)	11,700,000
Generator/turbine/switch gear (12,400 kw)	6,600,000
Wood handling system (104 tons per hour)	8,100,000
Cooling water (27,000 GPM)	2,760,000
Electrical system from main power grid (3,275 kw)	982,000
Boiler feedwater system (440 GPM)	490,000
Pollution control	
Pond and gypstack (Section 1100)	2,560,000
Sanitary waste (0.5% ISBL less CO_2 system)	300,000
Buildings (3% ISBL less CO_2 system)	1,815,000
General utilities (5% ISBL less CO_2 system)	3,024,000
Site development (6% ISBL less CO_2 system)	3,630,000
Piping (3% ISBL less CO_2 system)	1,815,000
	45,704,000

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TABLE VI-A-3

COST OF PRODUCTION ESTIMATE FOR ETHANOL
PROCESS- ENZYME HYDROLYSIS

CAPITAL SUMMARY

BASIS

Location: Michigan
Mid-1984

Capacity: 25.00 million gallons/yr

Str.Time: 8000 hours per year

CAPITAL COST	$MILLION
Battery Limits	66.3
Offsites	45.7
Total Fixed Inv.	112.0
Working Capital	8.7

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL
Mixed Hardwood, Lb	66.88212	.9	15,046	
Sulfuric Acid, Lb	2.23180	2.8	1,562	
Calcium Hydroxide, Lb	1.66251	1.6	665	
Sodium Hydroxide, Lb	.24741	7.5	464	
Catalyst & Chemicals			2,503	
TOTAL RAW MATERIALS			20,239	80.97
UTILITIES				
Purchased Power, kWH	.22404	6.0	336	
Generated Power, kWH	3.96878	1.3	1,290	
Bfw, M Gal	.00641	139.0	223	
Cooling Water, M Gal	.43609	9.8	1,068	
Steam,250 psia, M Lb	.08612	80.0	1,722	
Steam,600 psia, M Lb	.01480	90.0	333	
TOTAL UTILITIES			4,972	19.89
OPERATING COSTS				
Labor, 46 Men @ $ 27,900		10 M/S	1,283	
Foremen, 9 Men @ $ 31,700		1 M/S	285	
Supervision, 3 Man @ $ 38,200		3 Man	115	
Maint., Material & Labor		6% of ISBL	3,978	
TOTAL OPERATING COST			5,661	22.65
OVERHEAD EXPENSES				
Direct Overhead	45% Lab. & Sup.		757	
Gen. Plant Overhead	65% Oper. Costs		3,680	
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,680	
TOTAL OVERHEAD EXPENSES			6,117	24.47
BY-PRODUCT CREDIT				
Carbon Dioxide, Lb	6.51328	2.8	-4,558	
Furfural, Lb	1.46013	30.0	-10,949	
TOTAL BY-PRODUCT CREDIT			-15,507	-62.04
CASH COST OF PRODUCTION			21,482	85.95
DEPRECIATION	20% ISBL + 10% OSBL		17,830	71.33
NET COST OF PRODUCTION			39,312	157.28
REQUIRED SALES PRICE AT 10% DCF				206.5

Raw material costs total 80.97 cents per gallon of ethanol produced hardwood priced on at 18 dollars per wet ton, sulfuric acid at 56 dollars per ton and calcium hydroxide and sodium hydroxide at 32 dollars and 150 dollars per ton, respectively. A breakdown of catalyst and chemicals costs is presented in Table VI-A-4.

TABLE VI-A-4

CATALYST AND CHEMICALS SUMMARY

	Unit Cost ($/lb)	Lbs/Yr	Annual Cost, $
Cellobiase (65% moisture)	0.700	1,928,000	1,349,600
Sodium carbonate	0.062	3,024,000	187,500
Corn steep liquor (46% moisture)	0.081	11,784,000	954,500
Benzene	0.210	56,000	11,200
			2,502,800

Total utility costs amount to 19.89 cents per gallon of product. The operating and overhead expenses, both largely capital related, contribute 22.65 and 24.47 cents per gallon to the cost of ethanol.

In this facility the commercial by-products are furfural and carbon dioxide with by-product prices of 2.8 and 30.0 cents per pound for carbon dioxide and furfural respectively, there results a credit of 62.04 cents per gallon of ethanol.

The cash cost of production is 85.94 cents per gallon. The net cost of production, which includes depreciation, is 157.28 cents per gallon of product. Depreciation is calculated on a 5 and 10-year straight line basis for ISBL and offsites capital investments, respectively, accounting for 71.33 cents per gallon. In order to achieve a 10 percent discounted cash flow rate of return, an ethanol sales price of 206 cents per gallon is required.

3. Cash Flow Analysis

a. 100 Percent Equity Financing

A discounted cash flow (DCF) financial model has been developed for the ethanol facility project economic analysis. Table VI-A-5 presents the model summary for the base case design. The capital and production cost estimates presented above serve as the basis for the model.

Parameters used in the cash flow analysis are defined below:

- Project life - the project has a 15-year operating life following a 3-year construction period starting in 1984 and zero salvage value at the end of project life.

- Land cost - for this plant, it is estimated that the Michigan site consisting of 130 acres would cost 0.1 million dollars in 1984.

- Fixed investment cost - the total capital investment cost in 1984 dollars was presented earlier. A 3-year construction period, 1984 through 1986, is assumed.

- Start-up cost - these include provision for the expenses of a technical start-up team, consumption of feedstock and utilities, and other operating costs during this period of little or no real production. These costs are estimated at 10 percent of ISBL capital cost and are expended in the year of plant start-up (i.e., the last year of plant construction). These costs are amortized on a 5-year straight line basis and deducted from operating profit prior to the calculation of income taxes. These are included in the depreciation expense.

- Working capital - working capital requirements for the project are calculated using the procedure presented below:

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TABLE - VI- A-5
ENZYME HYDROLYSIS-100% EQUITY
ETHANOL FROM CELLULOSE

DISKETTE 262

	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
CAPITAL EXPENDITURES										
LAND	-.1	.0	.0	.0	.0	.0	.0	.0	.0	.0
FIXED INVESTMENT	-39.6	-39.6	-39.6	.0	.0	.0	.0	.0	.0	.0
START UP COST	.0	.0	-7.0	.0	.0	.0	.0	.0	.0	.0
WORKING CAPITAL	.0	.0	.0	-8.9	-2.4	-2.7	-.8	-.8	-.8	-.8
INTEREST DURING CONS	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
TOTAL CAPITAL EXPEND	-39.7	-39.6	-46.6	-8.9	-2.4	-2.7	-.8	-.8	-.8	-.8
DEBT	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
EQUITY	-39.7	-39.6	-46.6	-8.9	-2.4	-2.7	-.8	-.8	-.8	-.8
OPERATING RATE, %	.0	.0	.0	60.0	80.0	100.0	100.0	100.0	100.0	100.0
REVENUE	.0	.0	.0	40.8	56.5	73.3	76.1	78.6	81.3	84.0
EXPENSES										
WOOD	.0	.0	.0	-10.4	-14.6	-19.2	-20.1	-21.0	-22.0	-23.0
CHEMICALS	.0	.0	.0	-3.6	-5.0	-6.6	-6.9	-7.3	-7.6	-7.9
UTILITIES	.0	.0	.0	-3.7	-5.3	-7.2	-7.7	-8.2	-8.8	-9.4
LABOR	.0	.0	.0	-1.1	-1.6	-2.1	-2.2	-2.3	-2.4	-2.5
MAINTENANCE	.0	.0	.0	-4.7	-5.0	-5.3	-5.7	-6.0	-6.3	-6.7
DIRECT OVERHEAD	.0	.0	.0	-.9	-.9	-1.0	-1.0	-1.1	-1.1	-1.2
GENERAL PLANT OVHD.	.0	.0	.0	-4.4	-4.7	-5.0	-5.3	-5.6	-5.9	-6.2
INSURANCE	.0	.0	.0	-2.0	-2.1	-2.3	-2.4	-2.5	-2.7	-2.8
TOTAL EXPENSES	.0	.0	.0	-30.9	-39.4	-48.6	-51.4	-54.1	-56.9	-59.8
PROFIT, TAX AND CASH FLOW										
OPERATING MARGIN	.0	.0	.0	9.9	17.1	24.7	24.6	24.5	24.4	24.2
BSAR EXPENSES	.0	.0	.0	-2.4	-3.4	-4.4	-4.6	-4.7	-4.9	-5.0
DEPRECIATION	.0	.0	.0	-20.3	-20.3	-20.3	-20.3	-20.3	-4.9	-4.9
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
GROSS PROFIT	.0	.0	.0	-12.9	-6.6	.0	-.3	-.5	14.7	14.3
INTEREST PAYMENT	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
ADJUSTED PROFIT	.0	.0	.0	-12.9	-6.6	.0	-.3	-.5	14.7	14.3
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0	-14.7	-5.6
TAXABLE PROFIT	.0	.0	.0	-12.9	-6.6	.0	-.3	-.5	.0	8.8
CORPORATE TAX @ 45 %	.0	.0	.0	.0	.0	.0	.0	.0	.0	-3.9
INVESTMENT TAX CRED.	.0	.0	.0	.0	.0	.0	.0	.0	.0	3.9
NET PROFIT	.0	.0	.0	-12.9	-6.6	.0	-.3	-.5	.0	8.8
LOAN REPAYMENT	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
DEPRECIATION	.0	.0	.0	20.3	20.3	20.3	20.3	20.3	4.9	4.9
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0	14.7	5.6
CASH FLOW, 100% EQUITY	-39.7	-39.6	-46.6	-1.4	11.3	17.6	19.3	19.0	18.7	18.3

DCF IRR 5.9%

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TABLE - VI A-5
ENZYME HYDROLYSIS 100% EQUITY
ETHANOL FROM CELLULOSE

	1994	1995	1996	1997	1998	1999	2000	2001
CAPITAL EXPENDITURES								
LAND	.0	.0	.0	.0	.0	.0	.0	.0
FIXED INVESTMENT	.0	.0	.0	.0	.0	.0	.0	.0
START-UP COST	.0	.0	.0	.0	.0	.0	.0	.0
WORKING CAPITAL	-.9	-.9	-1.0	-1.0	-1.1	1.1	1.2	24.4
INTEREST DURING CONS	.0	.0	.0	.0	.0	.0	.0	.0
TOTAL CAPITAL EXPEND	.9	.9	-1.0	1.0	1.1	-1.1	1.2	24.4
DEBT	.0	.0	.0	.0	.0	.0	.0	.0
EQUITY	.9	.9	1.0	-1.0	1.1	-1.1	1.2	24.4
******								
OPERATING RATE, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
REVENUE	86.9	89.0	92.9	96.1	99.5	103.0	106.6	110.3
EXPENSES:								
WOOD	-24.0	-25.1	-26.2	-27.4	-28.6	-29.9	-31.3	-32.7
CHEMICALS	-8.3	-8.7	-9.0	-9.5	-9.9	10.3	10.8	11.3
UTILITIES	-10.1	-10.8	11.5	-12.3	-13.1	14.0	-15.0	16.0
LABOR	-2.6	-2.7	-2.9	-3.0	-3.1	3.3	-3.4	3.6
MAINTENANCE	-7.0	-7.4	-7.9	-8.3	-8.7	-9.2	-9.7	10.3
DIRECT OVERHEAD	-1.2	-1.3	-1.4	-1.4	-1.5	-1.5	-1.6	-1.7
GENERAL PLANT OVHD.	6.6	-7.0	-7.3	-7.8	-8.2	-8.6	-9.1	-9.6
INSURANCE	-3.0	-3.2	-3.3	-3.5	-3.7	-3.9	-4.1	4.4
TOTAL EXPENSES	-62.9	-66.1	-69.5	-73.1	-76.9	-80.9	-85.1	89.6
PROFIT,TAX AND CASH FLOW******								
OPERATING MARGIN	24.0	23.7	23.4	23.0	22.6	22.0	21.5	20.8
MSA&R EXPENSES	-5.2	-5.4	-5.6	-5.8	-6.0	6.2	-6.4	6.6
DEPRECIATION	-4.9	-4.9	-4.9	.0	.0	.0	.0	.0
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0
GROSS PROFIT	13.9	13.5	13.0	17.2	16.6	15.9	15.1	14.1
INTEREST PAYMENT	.0	.0	.0	.0	.0	.0	.0	.0
ADJUSTED PROFIT	13.9	13.5	13.0	17.2	16.6	15.9	15.1	14.1
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0
TAXABLE PROFIT	13.9	13.5	13.0	17.2	16.6	15.9	15.1	14.1
CORPORATE TAX @ 45 %	-6.3	-6.1	-5.8	-7.8	7.5	-7.1	6.8	-6.4
INVESTMENT TAX CRED.	6.3	2.4	.0	.0	.0	.0	.0	.0
NET PROFIT	13.9	9.8	7.1	9.5	9.1	8.7	8.3	7.8
LOAN REPAYMENT	0	.0	.0	.0	.0	.0	.0	.0
DEPRECIATION	4.9	4.9	4.9	.0	.0	.0	.0	.0
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0
******								
CASH FLOW,100% EQUITY	17.9	13.7	11.0	8.5	[illegible]	[illegible]	7.1	[illegible]
******								
DCF IRR	5.9%							

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- Feedstock inventory - 2 months supply of delivered chips plus 1/2 month's supply of the other raw materials valued at delivered prices.

- Finished product inventory - 1/2 month's supply of principal products and by-products valued at gross cost of production.

- Accounts receivable - 1 month's gross cost of production.

- Cash - 1 week's out-of-pocket expenses estimated at gross cost of production less depreciation.

- Warehouse/spare parts inventory - 3 percent ISBL capital cost.

LESS

- Accounts payable - 1 month's supply of raw materials at delivered prices.

Because working capital requirements increase in each year of the project life due to the impact of inflation and production levels, the annual increases are added as a negative cash flow item in the analyses. The total working capital expended is recouped in the last year of the project as a positive cash flow.

- Interest during construction - the project is based on 100 percent equity financing. Therefore, there is no interest or loan repayment required in the cash flow model.

- Operating capacity buildup - new process plants do not immediately reach nameplate production at start up. To allow for this, Chem Systems has assumed a capacity buildup to nameplate levels of three years, with plant operating capacities at 60 percent of nameplate in the year of start up, 80 percent in the second year of operation and 100 percent in the third year of plant operation. Operating levels remain at 100 percent of nameplate for the remaining years of plant operation.

- Escalation - the discounted cash flow analysis is performed on a current dollar basis. It is therefore necessary to escalate the elements in the financial analysis to reflect escalation in capital costs, product revenues and operating costs. Table VI-A-6 presents Chem Systems' estimate of the escalation rates for general inflation, capital costs and utility prices for the period 1984 through 2001.

TABLE VI-A-6

ESCALATION RATES, 1984-2001
(Percent per Year)

Period	GNP Deflator	Capital Cost	Utility Costs
1984-1990	5.0	6.5	7.5
1990-2001	4.5	5.5	6.9

- Revenues - the revenues are received from sale of the ethanol product as well as carbon dioxide and furfural by-products. Projected ethanol prices are based on a combined federal/state tax incentive of nine cents per gallon for 10 percent ethanol blends in gasoline. The price forecasts over the life of the project are presented in Section V.

- Expenses - the expenses correspond to those presented in the cost of production estimate for 1984, Table VI-A-3. For the years in the project life the expense elements in the financial analysis are escalated as presented in Table VI-A-7.

TABLE VI-A-7

ESCALATION RATES FOR EXPENSES

Wood	GNP
Chemicals	GNP
Utilities	Utilities
Labor	GNP
Maintenance	Capital
Direct overhead	GNP
General plant overhead	Capital
Insurance	Capital

- Operating margin - this is the revenues less operating expenses.

- MSARD Expense - this is the cost of management, sales, administration, research and development for the operation of the plant, taken at six percent of the total revenues.

- Depreciation - this is a non-cash cost and is based on straight line methods over five years for ISBL capital and ten years for offsite capital. The amortization of start-up expenses over five years is also included.

- IDC Amortization - interest during construction (IDC) does not apply.

- Gross profit - this is the operating margin less MSARD and depreciation expenses. It is the same as the adjusted profit since there is no interest expense.

- Loss carryforward - the losses for each operating year are carried forward and credited against tax liabilities in future years. In this project, the loses would be taken against the tax liabilities of 1992-1993.

- Taxable profit - this is the adjusted profit before tax less losses.

- Corporate tax - corporate tax is taken at 45 percent of profit. Losses are allowed to be carried forward at 100 percent toward the next years taxable income. Unused tax credits are carried forward until depletion.

- Investment tax credit - the investment tax credit is 10 percent of the total capital expenditure for the plant and is credited against current or future tax liabilities. In this case the credit is used in 1993 through the year 1995.

- Net profit - this is the after-tax profit, but is a non-cash profit since depreciation is included as an expense.

- Cash flow - this is the net profit plus depreciation.

- IRR - the internal rate of return (IRR) for this project (based on 100 percent equity financing) is 5.9 percent.

b. Debt Financing

An alternative to 100 percent equity financing in the DCF cash flow analysis is financing of the investment capital including land, fixed investment, start up and initial working capital based on 30 percent equity and 70 percent debt. This analysis is presented in Table VI-A-8. All the data are the same as the 100 percent equity case presented above with the exception of financing which involves interest and loan repayments. In this project the average interest rate is taken at 13 percent per annum and the loan repayment commences in the first year of operations. In the construction years the interest is capitalized (financed with debt and equity) and amortized in the first five years of plant operations.

With debt/equity financing the IRR on equity is negative for this project, indicating the high cost of debt financing relative to the present worth of the potential profits.

c. SERI's Comparative Case

In order to obtain a cash flow analysis which can be compared with economic analyses presented by other contractors, SERI has provided a set of economic parameters (Table VI-A-9) to be used for a standard cash flow analysis. Using these values, a base year 1984 cost of production estimate was calculated (Table VI-A-10). The cash flow analysis for 1984 through 2001 is presented in Table VI-A-11. In this analysis the ethanol selling price was calculated such that a 15 percent DCF return on equity

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TABLE - VI-A-8
ENZYME HYDROLYSIS-30% EQUITY
ETHANOL FROM CELLULOSE

DISKETTE - 262

	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
CAPITAL EXPENDITURES										
LAND	¯.1	.0	.0	.0	.0	.0	.0	.0	.0	.0
FIXED INVESTMENT	¯39.6	¯39.6	¯39.6	.0	.0	.0	.0	.0	.0	.0
START-UP COST	.0	.0	¯7.0	.0	.0	.0	.0	.0	.0	.0
WORKING CAPITAL	.0	.0	.0	¯8.9	¯2.4	¯2.7	¯.8	¯.8	¯.8	¯.8
INTEREST DURING CONS	¯1.9	¯5.9	¯10.5	.0	.0	.0	.0	.0	.0	.0
TOTAL CAPITAL EXPEND	¯41.6	¯45.5	¯57.1	¯8.9	¯2.4	¯2.7	¯.8	¯.8	¯.8	¯.8
DEBT	¯29.1	¯31.8	¯40.0	¯6.2	.0	.0	.0	.0	.0	.0
EQUITY	¯12.5	¯13.6	¯17.1	¯2.7	¯2.4	¯2.7	¯.8	¯.8	¯.8	¯.8
OPERATING RATE, %	.0	.0	.0	60.0	80.0	100.0	100.0	100.0	100.0	100.0
REVENUE	.0	.0	.0	40.8	56.5	73.3	76.1	78.6	81.3	84.0
EXPENSES										
WOOD	.0	.0	.0	¯10.4	¯14.6	¯19.2	¯20.1	¯21.0	¯22.0	¯23.0
CHEMICALS	.0	.0	.0	¯3.6	¯5.0	¯6.6	¯6.9	¯7.3	¯7.6	¯7.9
UTILITIES	.0	.0	.0	¯3.7	¯5.3	¯7.2	¯7.7	¯8.2	¯8.8	¯9.4
LABOR	.0	.0	.0	¯1.1	¯1.6	¯2.1	¯2.2	¯2.3	¯2.4	¯2.5
MAINTENANCE	.0	.0	.0	¯4.7	¯5.0	¯5.3	¯5.7	¯6.0	¯6.3	¯6.7
DIRECT OVERHEAD	.0	.0	.0	¯.9	¯.9	¯1.0	¯1.0	¯1.1	¯1.1	¯1.2
GENERAL PLANT OVHD.	.0	.0	.0	¯4.4	¯4.7	¯5.0	¯5.3	¯5.6	¯5.9	¯6.2
INSURANCE	.0	.0	.0	¯2.0	¯2.1	¯2.3	¯2.4	¯2.5	¯2.7	¯2.8
TOTAL EXPENSES	.0	.0	.0	¯30.9	¯39.4	¯48.6	¯51.4	¯54.1	¯56.9	¯59.8
PROFIT,TAX AND CASH FLOW										
OPERATING MARGIN	.0	.0	.0	9.9	17.1	24.7	24.6	24.5	24.4	24.2
MSARD EXPENSES	.0	.0	.0	¯2.4	¯3.4	¯4.4	¯4.6	¯4.7	¯4.9	¯5.0
DEPRECIATION	.0	.0	.0	¯20.3	¯20.3	¯20.3	¯20.3	¯20.3	¯4.9	¯4.9
IDC AMORTIZATION	.0	.0	.0	¯3.7	¯3.7	¯3.7	¯3.7	¯3.7	.0	.0
GROSS PROFIT	.0	.0	.0	¯16.5	¯10.2	¯3.7	¯3.9	¯4.2	14.7	14.3
INTEREST PAYMENT	.0	.0	.0	¯13.9	¯13.6	¯13.2	¯12.8	¯12.3	¯11.7	¯11.1
ADJUSTED PROFIT	.0	.0	.0	¯30.4	¯23.8	¯16.9	¯16.7	¯16.4	3.0	3.3
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0	¯3.0	¯3.3
TAXABLE PROFIT	.0	.0	.0	¯30.4	¯23.8	¯16.9	¯16.7	¯16.4	.0	.0
CORPORATE TAX @ 45 %	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
INVESTMENT TAX CRED.	.0	.0	.0	.0	.0	.0	.0	.0	.0	.0
NET PROFIT	.0	.0	.0	¯30.4	¯23.8	¯16.9	¯16.7	¯16.4	.0	.0
LOAN REPAYMENT	.0	.0	.0	¯2.7	¯3.0	¯3.4	¯3.8	¯4.3	¯4.9	5.5
DEPRECIATION	.0	.0	.0	20.3	20.3	20.3	20.3	20.3	4.9	4.9
IDC AMORTIZATION	.0	.0	.0	3.7	3.7	3.7	3.7	3.7	.0	.0
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0	3.0	3.3
30 % EQUITY	¯12.5	¯13.6	¯17.1	¯11.8	¯5.3	1.0	2.7	2.5	2.2	1.7
DCF IRR	¯6.0%									

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TABLE - VI A-8
ENZYME HYDROLYSIS-30% EQUITY
ETHANOL FROM CELLULOSE

	1994	1995	1996	1997	1998	1999	2000	2001
CAPITAL EXPENDITURES								
LAND	.0	.0	.0	.0	.0	.0	.0	.0
FIXED INVESTMENT	.0	.0	.0	.0	.0	.0	.0	.0
START-UP COST	.0	.0	.0	.0	.0	.0	.0	.0
WORKING CAPITAL	¯.9	¯.9	¯1.0	¯1.0	¯1.1	¯1.1	¯1.2	24.4
INTEREST DURING CONS	.0	.0	.0	.0	.0	.0	.0	.0
TOTAL CAPITAL EXPEND	¯.9	¯.9	¯1.0	¯1.0	¯1.1	¯1.1	¯1.2	24.4
DEBT	.0	.0	.0	.0	.0	.0	.0	.0
EQUITY	¯.9	¯.9	¯1.0	¯1.0	¯1.1	¯1.1	¯1.2	24.4
OPERATING RATE, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
REVENUE	86.9	89.8	92.9	96.1	99.5	103.0	106.6	110.3
EXPENSES								
WOOD	¯24.0	¯25.1	¯26.2	¯27.4	¯28.6	¯29.9	¯31.3	¯32.7
CHEMICALS	¯8.3	¯8.7	¯9.0	¯9.5	¯9.9	¯10.3	¯10.8	¯11.3
UTILITIES	¯10.1	¯10.8	¯11.5	¯12.3	¯13.1	¯14.0	¯15.0	¯16.0
LABOR	¯2.6	¯2.7	¯2.9	¯3.0	¯3.1	¯3.3	¯3.4	¯3.6
MAINTENANCE	¯7.0	¯7.4	¯7.9	¯8.3	¯8.7	¯9.2	¯9.7	¯10.3
DIRECT OVERHEAD	¯1.2	¯1.3	¯1.4	¯1.4	¯1.5	¯1.5	¯1.6	¯1.7
GENERAL PLANT OVHD.	¯6.6	¯7.0	¯7.3	¯7.8	¯8.2	¯8.6	¯9.1	¯9.6
INSURANCE	¯3.0	¯3.2	¯3.3	¯3.5	¯3.7	¯3.9	¯4.1	¯4.4
TOTAL EXPENSES	¯62.9	¯66.1	¯69.5	¯73.1	¯76.9	¯80.9	¯85.1	¯89.6
PROFIT, TAX AND CASH FLOW								
OPERATING MARGIN	24.0	23.7	23.4	23.0	22.6	22.0	21.5	20.8
MSARD EXPENSES	¯5.2	¯5.4	¯5.6	¯5.8	¯6.0	¯6.2	¯6.4	¯6.6
DEPRECIATION	¯4.9	¯4.9	¯4.9	.0	.0	.0	.0	.0
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0
GROSS PROFIT	13.9	13.5	13.0	17.2	16.6	15.9	15.1	14.1
INTEREST PAYMENT	¯10.3	¯9.5	¯8.6	¯7.6	¯6.4	¯5.1	¯3.6	¯1.9
ADJUSTED PROFIT	3.6	3.9	4.3	9.7	10.2	10.8	11.5	12.2
LOSS CARRYFORWD	¯3.6	¯3.9	¯4.3	¯9.7	¯10.2	¯10.8	¯11.5	¯12.2
TAXABLE PROFIT	.0	.0	.0	.0	.0	.0	.0	.0
CORPORATE TAX @ 45 %	.0	.0	.0	.0	.0	.0	.0	.0
INVESTMENT TAX CRED.	.0	.0	.0	.0	.0	.0	.0	.0
NET PROFIT	.0	.0	.0	.0	.0	.0	.0	.0
LOAN REPAYMENT	¯6.2	¯7.0	¯8.0	¯9.0	10.2	11.5	¯13.0	14.7
DEPRECIATION	4.9	4.9	4.9	.0	.0	.0	.0	.0
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0
LOSS CARRYFORWD	3.6	3.9	4.3	9.7	10.2	10.8	11.5	12.2
30 % EQUITY	1.3	.0	.3	¯.4	1.1	1.9	¯2.7	¯22.0
DCF IRR	-6.0%							

is realized over the life of the project. Assuming a 8.5 percent annual escalation of ethanol, the year-by-year ethanol selling prices required to achieve the 15 percent return are summarized in Table VI-A-12. Since the data used in the analysis do not coincide with Chem Systems forecast, specific details of the analysis are not discussed further.

TABLE VI-A-9

ECONOMIC ANALYSIS DATA
PROVIDED BY SERI

Capacity factor (hours/year)	8000
Energy costs (per KWh)	.06
Electricity buyback rate (per kWh)	.05
Inflation rate (%)	7
Base year for constant dollars	1984
Year of cost information	1984
Year of first commercial operation (3-yr. constr. period)	1987
Linear distribution of costs during construction period	
Facility lifetime (years)	15
Depreciation tax life (years)	5
ACRS depreciation schedule	15%,22%,21%,21%,21% per year
Salvage value	0
Taxes, insurance (each)	1% of investment
Income tax rate (%)	45
Investment tax credit (%)	10
Ratio of debt to capitalization (%)	30
Ratio of equity to capitalization (%)	70
Annual rate of return on debt (%)	13
DCF rate of return on equity (%)	15
Real escalation rate on capital costs (%)	1
Real escalation rate on operating costs (%)	1.5
Feedstock cost ($/dry ton)	42
Feedstock escalation rate (%)	1.5

B. Optimization Alternatives

In order to study variations of the base case design, a version of the enzyme hydrolysis simulation developed for SERI by Chem Systems(13) was modified to closely approximate this design. This resulted in an ethanol sales price on a mid-1984 basis of 206 cents per gallon using standard economic parameters.(13)

TABLE VI-A-10

COST OF PRODUCTION ESTIMATE FOR ETHANOL - SERI PROCESS- ENZYME HYDROLYSIS

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	Michigan	Battery Limits	66.3
	Mid-1984	Offsites	45.7
Capacity:	25.00 million gallons/yr		
Str.Time:	8000 hours per year	Total Fixed Inv.	112.0
		Working Capital	9.6

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL
Mixed Hardwood, Lb	66.88212	1.0	17,553	
Sulfuric Acid, Lb	2.23180	2.8	1,562	
Calcium Hydroxide, Lb	1.66251	1.6	665	
Sodium Hydroxide, Lb	.24741	7.5	464	
Catalyst & Chemicals			2,503	
TOTAL RAW MATERIALS			22,747	91.00
UTILITIES				
Purchased Power, kWH	.22404	6.0	336	
Generated Power, kWH	3.96878	1.3	1,290	
Bfw, M Gal	.00641	139.0	223	
Cooling Water, M Gal	.43609	9.8	1,068	
Steam,250 psia, M Lb	.08612	80.0	1,722	
Steam,600 psia, M Lb	.01480	90.0	333	
TOTAL UTILITIES			4,972	19.89
OPERATING COSTS				
Labor, 46 Men @ $ 27,900		10 M/S	1,283	
Foremen, 9 Men @ $ 31,700		1 M/S	285	
Supervision, 3 Man @ $ 38,200		3 Man	115	
Maint., Material & Labor		6% of ISBL	3,978	
TOTAL OPERATING COST			5,661	22.65
OVERHEAD EXPENSES				
Direct Overhead	45% Lab. & Sup.		757	
Gen. Plant Overhead	65% Oper. Costs		3,680	
Insurance, Prop. Tax	2.0% Tot. Fix. Inv.		2,240	
TOTAL OVERHEAD EXPENSES			6,677	26.71
BY-PRODUCT CREDIT				
Carbon Dioxide, Lb	6.51328	2.8	-4,558	
Furfural, Lb	1.46013	30.0	-10,949	
TOTAL BY-PRODUCT CREDIT			-15,507	-62.04
CASH COST OF PRODUCTION			24,550	98.22
DEPRECIATION	20% ISBL + 20% OSBL		22,400	89.62
NET COST OF PRODUCTION			46,950	187.84
REQUIRED SALES PRICE AT 10% DCF				215.5

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TABLE - VI-A-11
ENZYME HYDROLYSIS
ETHANOL FROM CELLULOSE - SERI ECONOMIC DATA

	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
CAPITAL EXPENDITURES										
LAND	-.1	.0	.0	.0	.0	.0	.0	.0	.0	.0
FIXED INVESTMENT	-40.3	40.3	-40.3	.0	.0	.0	.0	.0	.0	.0
START-UP COST	.0	.0	-7.2	.0	.0	.0	.0	.0	.0	.0
WORKING CAPITAL	.0	.0	.0	-14.4	-1.2	-1.3	-1.4	-1.5	-1.7	-1.8
INTEREST DURING CONS	-.0	-2.4	-4.3	.0	.0	.0	.0	.0	.0	.0
TOTAL CAPITAL EXPEND	-41.2	-42.7	-51.0	-14.4	-1.2	-1.3	-1.4	-1.5	-1.7	-1.8
DEBT	-12.4	-12.8	-15.5	-4.3	.0	.0	.0	.0	.0	.0
EQUITY	-28.8	-29.9	-36.2	-10.1	-1.2	-1.3	-1.4	-1.5	1.7	-1.8
OPERATING RATE, %	.0	.0	.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
REVENUE	.0	.0	.0	76.5	83.1	90.1	97.7	106.0	115.1	124.8
EXPENSES										
WOOD	.0	.0	.0	-22.4	-24.3	-26.4	-28.6	-31.1	-33.7	-36.6
CHEMICALS	.0	.0	.0	-6.6	-7.2	-7.8	-8.4	-9.1	-9.9	-10.8
UTILITIES	.0	.0	.0	-6.4	-6.9	-7.5	-8.2	-8.9	9.6	-10.4
LABOR	.0	.0	.0	-2.1	-2.3	-2.5	-2.7	-2.9	-3.2	-3.4
MAINTENANCE	.0	.0	.0	-5.0	-5.4	-5.8	-6.3	-6.8	-7.3	-7.9
DIRECT OVERHEAD	.0	.0	.0	-1.0	-1.1	-1.2	-1.3	-1.4	-1.5	-1.6
GENERAL PLANT OVHD.	.0	.0	.0	-4.6	-5.0	-5.4	-5.8	-6.3	-6.8	-7.3
INSURANCE	.0	.0	.0	-2.1	-2.3	-2.4	-2.6	-2.9	-3.1	-3.3
TOTAL EXPENSES	.0	.0	.0	-50.2	-54.4	-59.0	-63.9	-69.3	-75.1	-81.4
PROFIT, TAX AND CASH FLOW										
OPERATING MARGIN	.0	.0	.0	26.3	28.6	31.1	33.8	36.7	40.0	43.4
MSARD EXPENSES	.0	.0	.0	-4.6	-5.0	-5.4	-5.9	-6.4	6.9	-7.5
DEPRECIATION	.0	.0	.0	-19.2	-28.2	-26.9	-26.9	-26.9	.0	.0
IDC AMORTIZATION	.0	.0	.0	-1.5	-1.5	-1.5	-1.5	-1.5	.0	.0
GROSS PROFIT	.0	.0	.0	1.0	-6.1	-2.7	-.5	1.9	33.1	35.9
INTEREST PAYMENT	.0	.0	.0	-5.9	-5.7	-5.5	-5.4	-5.2	-4.9	-4.6
ADJUSTED PROFIT	.0	.0	.0	-4.9	-11.8	-8.3	-5.9	-3.2	20.1	31.3
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0	-20.1	-5.9
TAXABLE PROFIT	.0	.0	.0	-4.9	-11.8	-8.3	5.9	-3.2	.0	25.3
CORPORATE TAX @ 45 %	.0	.0	.0	.0	.0	.0	.0	.0	.0	11.4
INVESTMENT TAX CRED.	.0	.0	.0	.0	.0	.0	.0	.0	0	11.4
NET PROFIT	.0	.0	.0	-4.9	-11.8	-8.3	-5.9	-3.2	.0	25.3
LOAN REPAYMENT	.0	.0	.0	-1.1	-1.3	1.4	-1.6	-1.8	-2.1	-2.3
DEPRECIATION	.0	.0	.0	19.2	28.2	26.9	26.9	26.9	.0	.0
IDC AMORTIZATION	.0	.0	.0	1.5	1.5	1.5	1.5	1.5	.0	.0
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0	20.1	5.9
70 % EQUITY	-28.8	-29.9	-36.2	9.7	15.5	17.4	19.5	21.0	-24.4	27.2

DCF IRR 15.0%

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TABLE VI-A II
ENZYME HYDROLYSIS
ETHANOL FROM CELLULOSE - SERI ECONOMIC DATA

	1994	1995	1996	1997	1998	1999	2000	2001
CAPITAL EXPENDITURES								
LAND	.0	.0	.0	.0	.0	.0	.0	.0
FIXED INVESTMENT	.0	.0	.0	.0	.0	.0	.0	.0
START UP COST	.0	.0	.0	.0	.0	.0	0	.0
WORKING CAPITAL	2.0	2.1	2.3	2.5	2.7	2.9	3.2	41.0
INTEREST DURING CONS	.0	.0	.0	.0	.0	.0	.0	.0
TOTAL CAPITAL EXPEND	2.0	2.1	2.3	2.5	2.7	2.9	3.2	41.0
DEBT	.0	.0	.0	.0	.0	.0	.0	.0
EQUITY	2.0	2.1	2.3	2.5	2.7	2.9	3.2	41.0
OPERATING RATE, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
REVENUE	135.4	146.9	159.5	173.0	187.7	203.7	221.0	239.0
EXPENSES								
WOOD	39.7	43.1	46.7	50.7	55.0	59.7	64.7	70.3
CHEMICALS	11.7	12.7	13.8	14.9	16.2	17.6	19.1	20.7
UTILITIES	11.3	12.3	13.3	14.5	15.7	17.0	18.5	20.1
LABOR	3.7	4.0	4.4	4.7	5.2	5.6	6.1	6.6
MAINTENANCE	8.6	9.3	10.0	10.8	11.7	12.6	13.6	14.7
DIRECT OVERHEAD	1.8	1.9	2.1	2.3	2.5	2.7	2.9	3.1
GENERAL PLANT OVHD.	7.9	8.5	9.2	9.9	10.7	11.6	12.5	13.5
INSURANCE	3.6	3.9	4.2	4.5	4.9	5.3	5.7	6.2
TOTAL EXPENSES	88.2	95.6	103.7	112.4	121.8	132.0	143.1	155.1
PROFIT, TAX AND CASH FLOW								
OPERATING MARGIN	47.1	51.3	55.8	60.7	65.9	71.7	77.9	84.7
G&A EXPENSES	8.1	8.8	9.6	10.4	11.3	12.2	13.3	14.4
DEPRECIATION	.0	.0	.0	.0	.0	.0	.0	.0
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0
GROSS PROFIT	39.0	42.4	46.3	50.3	54.6	59.5	64.6	70.4
INTEREST PAYMENT	4.3	4.0	3.6	3.2	2.7	2.1	1.5	.0
ADJUSTED PROFIT	34.7	38.4	42.6	47.1	52.0	57.3	63.1	69.5
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0
TAXABLE PROFIT	34.7	38.4	42.6	47.1	52.0	57.3	63.1	69.5
CORPORATE TAX @ 45 %	15.6	17.3	19.2	21.2	23.4	25.8	28.4	31.3
INVESTMENT TAX CRED.	1.4	.0	.0	.0	.0	.0	0	.0
NET PROFIT	20.5	21.1	23.4	25.9	28.6	31.5	34.7	38.3
LOAN REPAYMENT	2.6	3.0	3.3	3.8	4.3	4.8	5.5	6.2
DEPRECIATION	.0	.0	.0	.0	.0	.0	.0	.0
IDC AMORTIZATION	.0	.0	.0	.0	.0	.0	.0	.0
LOSS CARRYFORWD	.0	.0	.0	.0	.0	.0	.0	.0
70 % EQUITY	15.9	16.1	17.0	19.6	21.6	23.8	25.1	23.1
DCF IRR	15.0%							

TABLE VI-A-12

REQUIRED PRICE OF ETHANOL TO ACHIEVE A 15 PERCENT DCF RETURN ON EQUITY

Year	Price, $/Gal
1984	1.78
1987	2.27
1988	2.46
1989	2.67
1990	2.90
1991	3.14
1992	3.41
1993	3.70
1994	4.01
1995	4.36
1996	4.73
1997	5.13
1998	5.56
1999	6.04
2000	6.55
2001	7.10

Two types of optimization studies were performed:

- Alternatives involving different equipment or processing steps using known and tested engineering techniques.

- Process improvements involving unproven but feasible advances in enzyme hydrolysis technology.

While the process design chosen as the basis for this study is only one of many that could have been chosen, it represents the current state of research in enzyme hydrolysis coupled with reasonable engineering assumptions. For the most part, optimization of processing steps involving standard commercially proven unit operations was carried out during the base case design effort and incorporated into the final process design. Some of these efforts are briefly discussed below.

Numerous alternatives to the commercially proven ethanol purification scheme used in this design have been proposed. Two such schemes were evaluate in detail:

- Multi-effect distillation (by Hoechst)

- Distillation with vapor recompression followed by potassium acetate extraction (by Dartmouth)

Both of these alternatives involved substantially less steam but more capital than the base case design. They were ruled out primarily to avoid substantial innovation in an area where conventional technology is adequate. It is reasonable to assume, however, that both alternatives would work and may offer some overall production cost savings if properly integrated into the process.

Multi-effect evaporation as employed in both sugar concentration and waste solids concentration involves a heavy capital investment and substantial energy usage. Mechanical recompression involving steam-driven compressors was investigated as an alternative for both six-effect evaporation systems. Recompression generally was found to be slightly more capital intensive, but required less steam. When considering the overall plant steam and power balances including power cogeneration from wood via letdown of high pressure steam, the conventional multi-effect evaporators were found to be the more economic alternative in both cases.

The option of eliminating sugar concentration entirely was investigated with the process simulator. This results in a reduction of the glucose concentration entering fermentation from 15 to 6.85 weight percent. While capital and utilities in this section of the plant decrease, there are substantial capital and utility cost increases in the fermentation, purification and heat generation sections associated with the greater water content present. This more than offsets the larger furfural by-product credit associated with a more dilute xylose stream entering furfural production. The net result is an increase of 17 cents per gallon in ethanol sales price as compared to the base case.

Another alternative investigated with the computer simulator was elimination of the waste treatment system. Since the condensate streams contain only 1,000 ppm of organic matter, it may be possible in certain circumstances to simply discharge this waste. Make-up process water would, however, be required. This alternative reduces the ethanol sales price by 2 cents per gallon, due to a slight decrease in capital which is somewhat offset by the addition of make-up process water requirements.

The waste streams currently routed to heat generation could be sent to an anaerobic digester where the organic content is reduced producing methane and carbon dioxide. Without recovery of the carbon dioxide from the anaerobic digester, this option is 6 cents per gallon more expensive than the base case. This is due to added capital for waste water treatment as well as additional raw materials to overcome the lower carbon utilization efficiency for steam generation. However, if the carbon dioxide is recovered along with the fermenter off-gas, the resultant ethanol sales price is 4 cents per gallon lower than the base case. This alternative includes a substantial increase in the carbon dioxide recovery system and its power requirements which partially offset the added by-product credit.

In summary, it seems that optimization of conventional technology areas offers little economic incentive as compared to the base case design. However, it is very important to determine which process parameters in the more innovative sections of the plant could be most effective in improving process economics. In this way the areas for future research and development can be established, which can ultimately result in an optimized enzyme hydrolysis process design.

Several different process improvements in pretreatment, enzyme production and enzyme hydrolysis were investigated using the simulation model. For each case a change was made in the design of one of the process subsections which is either based on a real research option or a hypothetical research goal. These cases are summarized below:

1. An enzyme recovery section was added in enzyme hydrolysis. It was assumed that by contacting the enzyme-sugar stream from enzyme hydrolysis with the solids feed to enzyme hydrolysis, that 60 percent of the original enzyme activity can be recovered. This is a conceptualized design, however, research at Berkeley[14] has shown that enzyme recovery may be feasible.

2. A steam digestor pretreatment replaced the prehydrolysis pretreatment of the base case. Recent research at Dartmouth[12] has indicated that a steam pretreatment at 350 psig and five-minute residence time may be as effective a pretreatment as mild acid prehydrolysis.

3. Enzyme productivity was increased from 52 IU/ml/hr (FPA 15 IU/ml) to 106 IU/ml/hr (FPA 30.4) which represents the best results obtained for solka floc in a fed-batch system. Also enzyme production per gram of carbon source consumed was increased from 0.17 to 0.31 grams enzyme per gram of carbon source. This is also the best result obtained for solka floc in a fed-batch system.

4. The solids concentration in enzyme hydrolysis was increased from 10 to 15 weight percent with no corresponding reduction in yield as predicted by experimentation, or increase in enzyme loading.

The first case, which incorporated enzyme recovery, reduced ethanol sales price by about 10 cents per gallon. This was due primarily to the reduction in raw materials cost by the 60 percent reduction in wood being diverted to enzyme production.

There was also a slight reduction in capital caused by the decrease in enzyme production tank size which was offset somewhat by the enzyme recovery equipment.

The second case, use of steam explosion in place of prehydrolysis for pretreatment, resulted in a four-cent per gallon ethanol cost reduction. This was primarily caused by the raw material cost reduction associated with the elimination of the acid and neutralizing chemicals added at the front end. There was a slight increase in capital due to the extra cost of disk refining at higher pressure and the increased solids (less xylan is solubilized during steam explosion compared to prehydrolysis) in the streams going to separation and enzyme hydrolysis.

The third case, increased enzyme productivity, resulted in a 17 cent per gallon ethanol sales price decrease. This is due to less feed diverted to enzyme production, resulting in decreases in wood, acid, calcium hydroxide and nutrient raw material requirements. Also capital in the front end is decreased due to the reduced wood feed, and enzyme production capital is reduced significantly resulting from the increased productivity.

The fourth case, increasing enzyme hydrolysis solids concentration to 15 weight percent, reduced ethanol cost 27 cents per gallon, the largest single decrease. Increasing the solids concentration in enzyme hydrolysis significantly reduces capital in enzyme hydrolysis and fermentation, (reduces size of sugar concentrator), as well as reducing the steam boiler cost in the offsites. The reduction in steam load also decreases the wood feed requirement and utilities cost.

A summary of the results for the alternative process schemes as compared to the base case at 10 percent DCF return are given in Table VI-B-1. If the results are assumed nearly cumulative and these are combined, the reasonable potential for reducing ethanol production costs via further process development can be analyzed. By combining these four cases a net decrease of approximately 50-55 cents per gallon could be realized. Since the base case ethanol sales price was 206 cents per gallon with a 10 percent DCF return, the optimized case might lie in the 150-160 cent per gallon range in mid-1984 dollars.

It must be emphasized that these optimization results represent attainable targets for future research and development. At the current state of enzyme technology development, even the 206 cent per gallon ethanol price would require experimental verification of design parameters.

Table VI-B-1

Summary of Process Alternatives

Alternative	Deviation from Base Case, Cents/Gal
1. Enzyme recovery	-10
2. Steam explosion	-4
3. Increased enzyme productivity	-17
4. 15% enzyme hydrolysis solids concentration	-27

VII. DISCUSSION

A. Sources and Limitations of Data

The bases for the design of the enzyme hydrolysis process described in this study are derived from many different sources. These include research efforts in biomass conversion technology, equipment manufacturers serving related areas such as the paper and pulp industry and Chem Systems experience in the biotechnology field. The heart of the process represents the fruits of the efforts of several researchers and design parameters are based on experimental results. Manufacturers of equipment for the paper and pulp and other industries have provided equipment specifications and costs for some of the larger package items such as the disk refiner, multi-effect evaporators and the wood boiler. Where no data exist or there is no analogous commercially available equipment, Chem Systems has extrapolated on existing data or made assumptions based on experience in biomass conversion processes. A brief discussion by plant section of the sources and limitations of the design and equipment bases ensues.

Section 100-Pretreatment/Prehydrolysis

The wood handling system equipment requirements and cost are based on information supplied by Charles T. Maine of Boston, Massachusetts, an engineering contractor who has designed and constructed a wood-based power plant for Burlington Electric in Vermont. A major concern with the wood handling system is whether a debarker is required. The main problem with bark is that it may carry dirt through the disk refiner, potentially damaging the disk grinders or hampering enzyme production and/or hydrolysis. Therefore the chips are thoroughly washed before entry into the disk refiner eliminating the need for the debarker.

The disk refiner is a commercially available item used in the paper and pulp industry for thermomechanical pulping. The equipment specifications and cost for a disk refiner producing particle fibers 0.5-1 millimeter in diameter were provided by C.E. Bauer of Springfield, Illinois, and Sprout Waldron of Munsey, Pennsylvania, both manufacturers of this type of equipment.

The prehydrolysis plug flow reactor (PFR) presented a unique problem since there is no commercially available equipment that is comparable. The design parameters and material balance around the reactor are based on Dartmouth experimental data as described in Section IV-B-1. However, there are some questions concerning minor components in the prehydrolysis product. The amount of acetic acid, other organic acids and phenolics formed during prehydrolysis has not been quantified. The presence of other organic acids could significantly affect the processing scheme in so much as these acids may have to be removed due to their potential toxic effects on the microbial systems downstream. In the very least they may have to be reneutralized. Chem Systems has made an assumption concerning the formation of actic acid based on the acetyl groups present as discussed in Section IV-B-1, however other byproducts have not been quantified.

The equipment conceptualized for the PFR has been based on the bench scale Dartmouth PFR and discussions with the disk refiner manufacturer, C.E. Bauer. The disk refiner is envisioned to serve as the feeder to the prehydrolysis reactor when operated at the PFR pressure of 230-240 psia. The PFR itself is similar to the Dartmouth pipe reactor. However, it has not been demonstrated that instantaneous heatup, uniform acid mixing and residence time control can be achieved for a scaled up reactor.

Section 200-Sugar Separation and Neutralization

The major concern in this section, and one of the principal concerns of the feasibility of the design in general is the solid-liquid separation characteristics of the materials being processed. Laboratory studies at Dartmouth have indicated that the filter rate of the prehydrolyzed ligno-cellulosic material is very low due to clogging of the filter. However, the material did settle out giving an indication that centrifuging would be possible. Experiments by Dartmouth using Bird centrifuges yielded filter cakes of 30-40 weight percent solids and demonstrated the feasibility of solid-liquid separations. However, these experiments have not been well documented, actual performance data are lacking, and certainly this separation has not been demonstrated on commercial sized equipment. Therefore a number of assumptions have been made concerning the performance of the solid-liquid separation equipment which are summarized in Section IV-B-2. Rotary drum filters are only used for cool (less than 70°C), low solids loading separations, such as polishing.

Section 300-Enzyme Production

Perhaps the largest question exists concerning the design parameters of the enzyme production system, since this section contributes significantly to the cost of production (approximately 30-40 cents per gallon). A great amount of research effort has been expended in this area, especially at Berkeley. However, design criteria for an optimized system are still lacking. Berkeley has individually demonstrated the feasibility of high enzyme concentrations and productivities for fed batch systems, and the use of low cost substrates (steam exploded wood) and nutrients (corn steep liquor) for enzyme growth. However, a single design incorporating all these advances into an optimized system has not been attempted. Therefore a number of assumptions have been made based on extrapolating what data are available. These assumptions have been summarized in Section IV-B-3. The major assumption made is that the carbon source used in this design, prehydrolyzed wood, will produce RUT C-30 enzymes at a productivity, concentration and carbon source consumption similar to steam exploded wood. In an attempt to verify this assumption, Chem Systems had Dartmouth conduct experiments on growing RUT C-30 on prehydrolyzed wood. The results of these experiments indicate that RUT C-30 will grow on the prehydrolyzed wood if it has been thoroughly washed, which apparently removes inhibitory effects of some prehydrolysis products. This is similar to results obtained by Berkeley.[8] However, the enzyme concentration and productivity are approximately half of what Berkeley obtained for steam exploded wood.[13] More research is needed in this area to verify these results.

Another area of concern in the enzyme production section (as well as the enzyme hydrolysis section) is the feasibility of agitation, pumpability and reaction of the 9-10 weight percent slurries in the enzyme production tanks. Goulds, a manufacturer of pumps, was contacted. The Goulds people felt that although testing of the material would have to be performed, based on their experience in pumping wood slurries, a recessed impeller type centrifugal pump would provide adequate agitation and pumpability of the material.

Section 400-Enzyme Hydrolysis

The yield in enzyme hydrolysis is based on experimental results obtained by Dartmouth (see Section IV-B-4) and is correlated to a number of design parameters including prehydrolysis conditions, enzyme loading, residence time and solids concentration. The yield assumed (i.e., 90 percent of theoretical) is considered to be somewhat conservative, since Dartmouth has obtained quantitative yields at the conditions used.

The cellobiase added to the RUT C-30 cellulase is a very small amount and is assumed to be purchased. Current price quotes from NOVO indicate that very small amounts of cellulase used for research purposes cost 5-7 dollars per pound. However, it is unlikely that bulk quantities of enzyme would be priced that high. In addition NOVO probably uses an expensive substrate like Solka Floc to grow their enzyme. In breaking out the cost of producing enzyme from wood in the current design a cost including return of 60-70 cents per pound of cellulase is obtained. Therefore a conservative value of 2 dollars per pound is used as a purchase price for cellobiase. Even if this value is too low, recent work at Dartmouth has indicated that the cellobiase addition level assumed in this study may be too high.

The equipment specifications and cost associated with the dewatering presses were obtained from Black Clawson of Middletown, Ohio. Based on discussions with Black Clawson, it was decided that it is reasonable to assume the dewatering presses could concentrate the ligno-cellulosic materials from 35 to 55 weight percent solids.

Section 500-Sugar Concentration

The six-effect forced circulation evaporator used for sugar concentration is a conventional item used in the chemical industry. The minimization of calcium sulfate plating is achieved by using an evaporator system with slurry recycle, which encourages calcium sulfate crystal growth as opposed to surface plating in the evaporator. The design, cost and equipment specification for the evaporator system were provided by the Unitech Division of Ecodyne, a manufacturer of various types of evaporator systems.

Section 600-Fermentation

Continuous sugar fermentation to ethanol is a well documented and commercially available technology. Ethanol yield and by-product formation are based on commercial fermentation ethanol data. By-products formed are limited for the sake of simplicity to glycerol, acetaldehyde, fusel oils, carbon dioxide and yeast. Fermentation residence time requirement is correlated with cell density for a continuous system with cell recycle and is based on industrial practice.

A major problem in fermentation, especially with yeast recycle is contamination. In order to minimize the possibility of contamination, the yeast being recycled is acid washed at low pH for several hours. In addition, in order to avoid mutations, complete yeast population change out is achieved within 1-2 weeks by purging the yeast grown during fermentation. Also a select yeast strain is chosen.

The design specifications of the yeast centrifuges were obtained from information from Alfa-Laval, of Ft. Lee, New Jersey. Alfa-Laval manufacters a wide range of equipment for the fermentation industry.

Section 700-Carbon Dioxide Recovery

The carbon dioxide recovery unit is a standard sized (300 ton per day) package unit. The design specifications and cost were provided by a major industrial gas supplier.

Section 800-Ethanol Purification

The design of the ethanol purification system is based on the cascade distillation system developed by Katzen,[14] and used in commercial fermentation ethanol facilities. One design alternative which warrants careful evaluation is the distillation/heat pump system followed by potassium acetate extraction of ethanol under development at Dartmouth. This scheme is potentially highly energy efficient at low ethanol concentrations. When coupled with the elimination of multi-effect evaporation prior to fermentation and the replacement of concentration in the heat generation section with anaerobic digestion, the net result may be more economic than the standard systems proposed in this study. Due to second-order reaction kinetics, the furfural yield would be greater from the more dilute solution encountered in this scheme. These interactions require investigation and, if sufficient economic potential exists, this ethanol purification system should be experimentally verified.

Section 900-Furfural Production

The material balance and design of the furfural production reactor are based on experimental data provided by Dartmouth, which is summarized in Section IV-B-9. The reactor itself is similar to the prehydrolysis PFR except that it is a liquid phase reactor, which eliminates many of the problems associated with the slurry reactor. The reactor is envisioned to be a column with baffle plates to simulate plug flow (i.e., minimize back-mixing). As with the prehydrolysis PFR, materials of construction present a problem in the high temperature acidic environment of the system. Zirconium has proved satisfactory in the Dartmouth PFR. Also, as with the prehydrolysis PFR, an exact material balance quantifying all by products is not available. The design described herein does not quantify any light components which may be formed (i.e., methanol). However, a column is included in the design which removes lights (ethanol) present in the feed.

Section 1000-Heat Generation

The heat generation section is a relatively straightforward design using commercially available equipment. The stillage evaporator is a six-effect forced circulation type. Design specifications and cost for the stillage evaporator were provided by the Unitech Division of Ecodyne. The wood boiler/cogeneration system which handles a combination wood/organic liquid feed is a standard package unit which is used in commercial wood-based electric generating stations. Design specifications and cost for this system were provided by CE Power Systems of New York City.

Section 1100-Waste Treatment

The waste treatment system handles all the waste water from the process and decreases the organic content from 1000 to 20 ppm by aerobic digestion. The design specifications and cost were provided by EIMCO of Salt Lake City, Utah. The equipment specified is of the type commonly used to reduce the organic content of the water effluent from fermentation ethanol plants.

General

A general comment concerning the equipment specified is that varying degrees of technical applicability exist. That is, some of the equipment specified is used commercially for very closely related applications. This includes the concentrators, waste ponds, wood boiler cogeneration, carbon dioxide recovery and the distillation and fermentation equipment. However, other equipment specified would have to be extensively tested with respect to their specific application in this design. This would include the centrifuges, disk refiner/prehydrolysis reactor, slurry pumps, dewatering presses and filters.

B. Critical Technology Issues

The major critical technology issue concerning the process described in this study can be succinctly summarized as the demonstration of several individual unit operations and the technical feasibility of several process steps in a scaled-up integrated facility. Although many of the process steps have been individually demonstrated in the laboratory, it has not been shown that similar results can be obtained for a feedstock undergoing several process steps in succession. This of course would require a pilot plant-sized demonstration facility including, at a minimum, pretreatment, separation, enzyme production, enzyme hydrolysis and fermentation processing steps. The fermentation of wood hydrolyzate via pretreatment and enzymatic hydrolysis has to be studied in a continuous system to determine the acceptable level of inhibitors or toxins from pretreatment as well as to demonstrate that hydrolysis enzymes and fermentation yeasts can adapt to certain toxin levels.

The individual unit operations which are questionable for this specific application have been discussed with regard to the basis and justification of their selection in the previous section. These include the centrifuges/filters/dewatering processes for solid-liquid separations, slurry pumps for high-solids transfer and agitation, and the disk refiner/prehydrolysis reactor for pretreatment. The feasibility of these items in performing the desired unit operations has yet to be demonstrated on commercial sized equipment. Only the prehydrolysis reactor (using a slurry pump instead of the disk refiner as the feeder) and some centrifuge experiments have been demonstrated on a bench scale. Testing of the equipment to verify suitability for these applications is required in collaboration with the various manufacturers. This involves supplying the vendors with suitable quantities of material for testing, which unfortunately is not readily available.

It has not yet been demonstrated that disk refined prehydrolyzed wood can actually be used for enzyme production, hydrolyzed to sugar using such enzymes and then fermented to ethanol in integrated operation. In addition, because of the incompleteness of the material balance and the speculation on the effect of by-product and salt formation it is unknown what processing steps may be required upstream and downstream in order to purify the stream of all toxic components prior to entry into the microbial systems. Enzyme production probably requires washing of the unneutralized solids to be used as the carbon source. Recent experiments at Dartmouth(13) have already indicated that some toxic materials are apparently present in prehydrolyzed wood inhibiting cell growth and enzyme production. Although it has been shown by these experiments that the presence of poplar bark apparently does not inhibit enzyme hydrolysis of prehydrolyzed wood, its effect on enzyme production is unknown. Also the effect of bark from other wood species and the by-products formed during prehydrolysis are unknown. Solutions to these problems involve the possibility of using some kind of steaming as a pretreatment, thus minimizing the by-product formation associated with mild acid hydrolysis. Overneutralization of the prehydrolyzate to precipitate some of the phenolic by-products formed during prehydrolysis may also facilitate their removal. These options are discussed in more detail in Section VII-D.

C. Commercial Potential

The commercial potential of the enzyme hydrolysis process described in this study, or, for that matter, any cellulose hydrolysis process, will be dependent on the demonstration of the technical feasibility of several of the process steps discussed in Section VII-B. If technical feasibility is established, then the economic competitiveness of the process, vis a vis corn and acid hydrolysis based ethanol, will determine its commercial potential.

The base case design for the enzyme hydrolysis process produced ethanol for 206 cents per gallon at 10 percent DCF. This facility produces 25 million gallons per year of ethanol in Michigan (in mid-1984 assuming instantaneous plant construction and start-up. A 50 million gallon per year whole kernal milling corn ethanol plant at a U.S. Midwest location produces ethanol for 219 cents per gallon at 10 percent DCF in mid-1984. Costs of production from these two plants are compared in Table VII-C-1.

The wood enzyme hydrolysis case produces ethanol for about 13 cents per gallon less than the corn ethanol case, based on a new plant with instantaneous construction and start-up in 1984. However, these estimates are based on a current corn price of 3.10 dollars per bushel, which is expected to be one of the highest corn prices for the years 1984-2001. Therefore, ethanol cost of production from the corn plant can be expected to be somewhat lower over the plant lifetime.

It is presumed that a wood based acid hydrolysis facility would produce ethanol for an even lower cost of production, perhaps 180 cents per gallon, than the enzyme or corn cases on a 1984 basis.

TABLE VII-C-1

COMPARISON OF ETHANOL PROCESS ECONOMICS

	Wood Enzyme Hydrolysis (25 MM Gal/Yr)	Corn Whole Kernel Milling (50 MM Gal/Yr)
Investment, MM $		
Battery Limits	66.3	80.6
Offsites	45.7	62.8
Total Fixed Investment	112.0	143.4
Cost of Production, Cents/Gal		
Raw Materials	80.97(1)	130.89(2)
Utilities	19.89	43.90
Operating Costs	22.65	12.89
Overhead Expenses	24.47	14.13
By-product Credit	(62.04)	(67.86)
Cash Cost of Production	85.95	133.96
Depreciation	71.33	44.83
Net Cost of Production	157.28	178.79
Selling price at 10% DCF	206.0	218.5

(1) Mixed hardwoods at $18/wet ton.
(2) Corn kernels at $3.10/bushel.

The commercial potential for the enzyme hydrolysis process is probably dependent on whether the potential improvements discussed in Section VI-B can be realized. The base enzyme case is certainly competitive with a corn ethanol facility on a current basis, however, significant process improvements are still necessary for the enzyme hydrolysis process to become competitive with a wood-based acid hydrolysis process. Furthermore, the technical feasibility of enzyme hydrolysis must be demonstrated in an integrated facility before it will gain acceptance in the commercial marketplace as a viable alternative to corn-based routes.

D. Recommendations for Future R&D

The most important goal of future research and development work for a wood-based enzyme hydrolysis ethanol process is the demonstration of the technical viability of several unit operations as well as the satisfactory performance of an integrated process. This has been discussed in terms of process sensitivities in Section VI-B. This would likely require the construction and operation of a pilot plant sized unit demonstrating at a minimum the pretreatment, separation, enzyme production, enzyme hydrolysis and fermentation process steps.

Bench scale research should concentrate on those areas where process economics can be improved most significantly. These areas include, as discussed in Section VI-B, optimization of pretreatment, enzyme hydrolysis and enzyme production. Additionally, bench scale research is necessary to characterized the, as yet, incomplete material balance with regard to tars and minor organics such as acetic acid, methanol, levulinic acid, formic acid and soluble phenolics from lignin breakdown. Understanding these minor constituents is a prerequisite to finalizing this process design.

The optimized pretreatment serves two important functions. The first and most important is to give high enzyme hydrolysis yields in the most cost effective manner, (low capital, low enzyme loading and low hydrolysis residence time). However, it is also important that the pretreated wood feedstock serve as an inexpensive, effective carbon source for enzyme production. So far it has not been demonstrated that prehydrolyzed wood can be used as a carbon source for enzyme production and yield enzyme productivities as high as steam exploded wood. However, although steam exploded wood yields satisfactory enzyme productivities, it has not been shown to give enzyme hydrolysis yields as high as prehydrolysis when supplemented by cellobiase. More research is needed to verify these preliminary results and to optimize the pretreatment system.

Enzyme production is another area where significant cost reductions can be achieved by research improvements. The ultimate research goal should be to be able to utilize an inexpensive substrate like steam exploded or prehydrolyzed wood and achieve similar enzyme yields and productivities to those achieved with solka floc. Also optimization of the enzyme production system including nutrient type and requirement, and system type (batch, fed-batch or continuous) is required. Partial recovery of enzymes following hydrolysis coupled with the above improvements would reduce enzyme production costs even further. This is another area where more research needs to be done.

As evidenced by the discussion in Section VI-B, the key to significantly improved process economics may be in the amount of water required in enzyme hydrolysis. An increase in solids concentration of 5 percent (increasing from 10 to 15 percent) in enzyme hydrolysis reduces costs of production by 27 cents per gallon if the same hydrolysis yield is assumed. At present, as solids concentration increases, hydroysis yields correspondingly decrease. It is not known exactly what causes this decrease in yield. However, it is presumed that mixing difficulties and product inhibition (glucose) are the major factors. This problem should be researched for possible improved mixing techniques and physical or chemical removal of the glucose as it is formed. Also, minimum enzyme loading for both RUT C-30 cellulase and cellobiase should be achieved.

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IX. ACKNOWLEDGEMENTS

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X. APPENDIX

A. Detailed Material Balance

This sub-section contains the detailed material balance for the eleven plant sections based on 25 million gallons per year of fuel-grade ethanol.

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 100: PRETREATMENT/PREHYDROLYSIS

STREAM NO.		101		102		103		104		105		106		107		108		109	
DISKETTE 420 FILE SEC1	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	92905.9	5161.4	48600.7	2700.0	141514.5	7861.9	303675.0	16870.8	0.0	0.0	63931.9	3551.8	509121.5	28284.5	506823.6	28156.9	58720.6	3262.3
CELLULOSE	162	43851.7	270.7	0.0	0.0	43851.7	270.7	0.0	0.0	0.0	0.0	0.0	0.0	43851.7	270.7	41601.6	256.8	0.0	0.0
INSOLUBLE LIGNIN	---	17094.6	---	0.0	---	17094.6	---	0.0	---	0.0	---	0.0	---	17094.6	---	15384.0	---	0.0	---
ASH	---	185.6	---	0.0	---	185.6	---	0.0	---	0.0	---	0.0	---	185.6	---	185.6	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	9848.2	---	0.0	---	9848.2	---	0.0	---	0.0	---	0.0	---	9848.2	---	9854.0	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	15058.7	114.0	0.0	0.0	15058.7	114.0	0.0	0.0	0.0	0.0	0.0	0.0	15058.7	114.0	4391.6	33.3	0.0	0.0
HEMI. GLUCAN	162	4273.4	26.4	0.0	0.0	4273.4	26.4	0.0	0.0	0.0	0.0	0.0	0.0	4273.4	26.4	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	2601.0	---	0.0	---	2601.0	---	0.0	---	0.0	---	0.0	---	2601.0	---	2601.0	---	0.0	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	1709.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	484.6	5.0	210.9	2.2
HMF	126	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	55.7	0.4	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	11333.2	75.6	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	7160.2	39.8	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3847.4	39.3	0.0	0.0	3847.4	39.3	3847.4	39.3	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1161.7	19.4	67.4	1.1
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	67.4	2.1	55.7	1.7
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		185811.7		48600.7		234420.4		303675.0		3847.4		63931.9		605874.0		605873.0		59054.7	
TEMPERATURE, C		18.0		204.0		200.0		27.0		27.0		204.0		200.0		200.0		148.0	
PRESSURE, ATM		1.0		17.0		16.0		1.0		1.0		17.0		15.3		15.3		4.4	

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 100: PRETREATMENT/PREHYDROLYSIS

STREAM NO.		110		111		112		113		114		115		116		117	
DISKETTE 420 FILE SEC1	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	440103.0	24894.6	28547.0	1585.9	419556.0	23308.7	15840.1	880.0	403715.9	22428.7	44387.1	2465.9	9806.3	544.8	48314.3	2717.5
CELLULOSE	162	41601.6	256.8	0.0	0.0	41601.6	256.8	0.0	0.0	41601.6	256.8	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	15384.0	---	0.0	---	15384.0	---	0.0	---	15384.0	---	0.0	---	0.0	---	0.0	---
ASH	---	185.6	---	0.0	---	185.6	---	0.0	---	185.6	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	9054.0	---	0.0	---	9054.0	---	0.0	---	9054.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	4393.6	33.3	0.0	0.0	4393.6	33.3	0.0	0.0	4393.6	33.3	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	2601.0	---	0.0	---	2601.0	---	0.0	---	2601.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	1709.0	---	0.0	---	1709.0	---	0.0	---	1709.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	273.7	2.9	90.2	0.9	183.5	1.9	45.7	0.5	137.8	1.4	135.9	1.4	35.2	0.4	175.7	1.8
HMF	126	55.7	0.4	0.0	0.0	55.7	0.4	0.0	0.0	55.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	11333.2	75.6	0.0	0.0	11333.2	75.6	0.0	0.0	11333.2	75.6	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	7168.2	39.8	0.0	0.0	7168.2	39.8	0.0	0.0	7168.2	39.8	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	3847.4	39.3	0.0	0.0	3847.4	39.3	0.0	0.0	3847.4	39.3	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	1094.2	18.2	36.1	0.6	1058.1	17.6	18.6	0.3	1039.5	17.3	54.7	0.9	11.3	0.2	56.2	0.9
METHANOL	32	11.7	0.4	11.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0	11.7	0.4	9.3	0.3	46.4	1.4
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		546819.2		28685.0		518134.2		15904.4		502229.8		44589.4		9862.1		49192.5	
TEMPERATURE, C		110.0		110.0		110.0		100.0		100.0		100.0		140.0		140.0	
PRESSURE, ATM		4.4		2.0		2.0		1.1		1.1		1.1		4.4		4.4	

DATE 18 Aug 84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 200: SUGAR SEPARATION/NEUTRALIZATION

STREAM NO.		201		202		203		204		205		206		207		208		209	
DISKETTE 420 FILE SECZ	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	403715.9	22428.7	116542.4	6474.6	287173.5	15954.1	5827.0	323.8	281345.7	15630.3	0.0	0.0	19480.4	1082.2	302030.8	16779.5	12918.9	717.7
CELLULOSE	162	41601.6	256.8	39521.6	244.0	2080.0	12.8	1975.5	12.2	104.5	0.6	0.0	0.0	0.0	0.0	104.5	0.6	104.5	0.6
INSOLUBLE LIGNIN	---	15384.0	---	14615.9	---	760.9	---	729.0	---	39.1	---	0.0	---	0.0	---	39.1	---	39.1	---
ASH	---	185.6	---	175.9	---	9.0	---	9.8	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	9054.0	---	8602.5	---	452.4	---	429.9	---	22.5	---	0.0	---	0.0	---	22.5	---	22.5	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3729.2	27.4	3261.2	24.0
HEMI. XYLAN	132	4391.6	33.3	4173.7	31.6	219.8	1.7	209.1	1.6	10.7	0.1	0.0	0.0	0.0	0.0	10.7	0.1	10.7	0.1
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	2601.0	---	751.3	---	1850.4	---	37.1	---	1813.3	---	0.0	---	0.0	---	1813.3	---	77.2	---
SOLUBLE LIGNIN	---	1709.0	---	493.4	---	1216.4	---	24.4	---	1191.9	---	0.0	---	0.0	---	1191.9	---	50.8	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	137.0	1.4	40.1	0.4	97.7	1.0	2.0	.0	95.7	1.0	0.0	0.0	0.0	0.0	95.7	1.0	3.9	.0
HMF	126	55.7	0.4	15.6	0.1	40.1	0.3	1.0	.0	39.1	0.3	0.0	0.0	0.0	0.0	39.1	0.3	2.0	.0
XYLOSE	150	11333.2	75.6	3272.0	21.8	8061.2	53.7	163.2	1.1	7890.1	52.7	0.0	0.0	0.0	0.0	7890.1	52.7	338.0	2.3
GLUCOSE	180	7160.2	39.8	2069.3	11.5	5099.0	28.3	103.6	0.6	4995.4	27.8	0.0	0.0	0.0	0.0	4995.4	27.8	214.0	1.2
SULFURIC ACID	98	3847.4	39.3	1110.0	11.3	2736.6	27.9	49.0	0.5	2686.0	27.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2476.7	33.5	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	954.5	6.0	41.0	0.3
ACETIC ACID	60	1039.5	17.3	299.9	5.0	739.6	12.3	14.7	0.2	724.9	12.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		502229.0		191684.5		310545.3		9577.5		300967.0		2476.7		19480.4		322924.9		17083.8	
TEMPERATURE, C		100.0		100.0		100.0		100.0		60.0		27.0		27.0		60.0		63.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

STREAM NO.		210		211		212		213		214		215		216	
DISKETTE 420 FILE SEC2	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	289111.9	16061.8	616.5	34.2	12302.4	683.5	301414.3	16745.2	122370.2	6798.3	14684.3	815.8	107685.9	5982.6
CELLULOSE	162	0.0	0.0	99.7	0.6	4.9	.0	4.9	.0	41497.1	256.2	4979.8	30.7	36517.3	225.4
INSOLUBLE LIGNIN	---	0.0	---	37.1	---	2.0	---	2.0	---	15345.7	---	1841.6	---	13504.1	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	185.6	---	22.5	---	163.2	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	21.5	---	1.0	---	1.0	---	9032.4	---	1083.5	---	7940.9	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	469.0	3.4	3079.5	22.6	101.7	1.3	650.7	4.8	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	9.8	0.1	1.0	.0	1.0	.0	4382.8	33.2	525.6	4.0	3857.2	29.2
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	1736.1	---	3.9	---	73.3	---	1809.4	---	788.4	---	94.8	---	693.7	---
SOLUBLE LIGNIN	---	1141.1	---	2.9	---	47.9	---	1189.0	---	517.0	---	62.5	---	455.3	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	91.8	1.0	0.0	0.0	3.9	.0	95.7	1.0	42.0	0.4	4.9	0.1	37.1	0.4
HMF	126	37.1	0.3	0.0	0.0	2.0	.0	39.1	0.3	16.6	0.1	2.0	.0	14.7	0.1
XYLOSE	150	7560.0	50.4	16.6	0.1	321.4	2.1	7881.5	52.5	3435.1	22.9	412.3	2.7	3022.8	20.2
GLUCOSE	180	4781.4	26.6	10.7	0.1	203.2	1.1	4984.7	27.7	2172.8	12.1	268.9	1.4	1912.0	10.6
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1160.7	11.8	139.7	1.4	1021.0	10.4
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	913.5	5.8	2.0	.0	39.1	0.2	952.6	6.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	314.6	5.2	38.1	0.6	276.5	4.6
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		305042.0		3900.2		13183.6		319025.7		201262.0		24152.4		177103.6	
TEMPERATURE, C		83.0		83.0		83.0		83.0		98.0		98.0		98.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

DATE 18-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 300: ENZYME PRODUCTION

STREAM NO.		301		302		303		304		305		306		307		308		309	
DISKETTE 420 FILE SEC3	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	14684.3	815.8	22329.3	1240.5	0.0	0.0	0.0	0.0	678.0	37.7	18649.0	1036.1	56403.2	3133.5	0.0	0.0	33865.0	1881.4
CELLULOSE	162	4979.0	30.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	1841.6	---	0.0	---	0.0	---	0.0	---	0.0	---	1841.6	---	3683.3	---	0.0	---	3479.6	---
ASH	---	22.5	---	0.0	---	0.0	---	0.0	---	0.0	---	22.5	---	44.9	---	0.0	---	43.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	340.0	---	697.6	---	0.0	---	662.4	---
OTHER INSOLUBLES	---	1083.5	---	0.0	---	0.0	---	0.0	---	0.0	---	1083.5	---	2167.0	---	0.0	---	2058.5	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	525.6	4.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	14107.5	440.9	0.0	0.0	0.0	0.0	0.0	0.0	12697.1	396.8	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	53072.6	1895.4	0.0	0.0	0.0	0.0	0.0	0.0	53072.6	1895.4	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	5774.4	131.2	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	795.3	---	11.7	---	807.0	---	0.0	---	484.6	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	13.7	---	1480.2	---	0.0	---	889.1	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	94.0	---	0.0	---	0.0	---	0.0	---	0.0	---	1.0	---	95.7	---	0.0	---	57.6	---
SOLUBLE LIGNIN	---	62.5	---	0.0	---	0.0	---	0.0	---	0.0	---	1.0	---	63.5	---	0.0	---	38.1	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	4.9	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	4.9	0.1	0.0	0.0	2.9	.0
HMF	126	2.0	.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.0	.0	0.0	0.0	1.0	.0
XYLOSE	150	412.3	2.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	260.9	1.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	139.7	1.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.4	.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	138.7	3.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.0	.0	204.2	1.4	0.0	0.0	123.1	0.9
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	0.0	52.0	0.6	0.0	0.0	31.3	0.4
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	38.1	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		24152.4		22329.3		138.7		67180.5		1473.3		21975.7		65706.6		71541.0		41757.0	
TEMPERATURE, C		100.0		27.0		27.0		27.0		27.0		27.0		28.0		28.0		28.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

STREAM NO.		310		311		312		313		314		315		316		317	
DISKETTE 420 FILE SEC3	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	22537.4	1252.1	20878.5	1159.9	54744.2	3041.3	33673.5	1870.7	78398.4	4355.5	57327.4	3184.9	100935.8	5607.5	15024.3	834.7
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	103.7	---	0.0	---	3499.6	---	3324.7	---	174.9	---	0.0	---	358.6	---	1483.1	---
ASH	---	2.0	---	0.0	---	43.0	---	41.0	---	2.0	---	0.0	---	3.9	---	18.6	---
MYCELLIUM	---	35.2	---	0.0	---	662.4	---	629.2	---	33.2	---	0.0	---	68.4	---	280.4	---
OTHER INSOLUBLES	---	108.4	---	0.0	---	2058.5	---	1956.0	---	102.6	---	0.0	---	211.0	---	872.5	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	322.4	---	0.0	---	484.6	---	14.7	---	469.9	---	0.0	---	792.3	---	2.9	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	591.1	---	0.0	---	889.1	---	26.4	---	862.7	---	0.0	---	1453.8	---	12.7	---
CELLODIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	38.1	---	0.0	---	57.6	---	2.0	---	55.7	---	0.0	---	93.8	---	1.0	---
SOLUBLE LIGNIN	---	25.4	---	0.0	---	38.1	---	1.0	---	37.1	---	0.0	---	62.5	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	2.0	.0	0.0	0.0	2.9	.0	0.0	0.0	2.9	.0	0.0	0.0	4.9	0.1	0.0	0.0
HMF	126	1.0	.0	0.0	0.0	1.0	.0	0.0	0.0	1.0	.0	0.0	0.0	2.0	.0	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	81.1	0.6	0.0	0.0	123.1	0.9	3.9	.0	119.2	0.8	0.0	0.0	200.3	1.4	2.0	.0
SODIUM ACETATE	82	21.5	0.3	0.0	0.0	31.3	0.4	1.0	.0	30.3	0.4	0.0	0.0	51.8	0.6	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		23949.2		20878.5		62635.5		39673.0		80289.9		57327.4		104239.1		17697.4	
TEMPERATURE, C		28.0		27.0		28.0		27.0		27.0		27.0		27.0		27.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

DATE 10-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS   SECTION 400: ENZYME HYDROLYSIS

STREAM NO.		401		402		403		404		405		406		407		408		409	
DISKETTE 450 FILE SEC4	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	107685.9	5982.6	301414.3	16745.2	0.0	0.0	26435.7	1468.6	100935.0	5607.5	156.3	0.7	537083.3	29838.0	532904.7	29605.8	38404.9	2133.6
CELLULOSE	162	36517.3	225.4	4.9	.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	36522.2	225.4	3652.0	22.5	0.0	0.0
INSOLUBLE LIGNIN	---	13504.1	---	2.0	---	0.0	---	0.0	---	358.6	---	0.0	---	13864.6	---	13864.6	---	0.0	---
ASH	---	163.2	---	0.0	---	0.0	---	0.0	---	3.9	---	0.0	---	167.1	---	167.1	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	68.4	---	0.0	---	68.4	---	68.4	---	0.0	---
OTHER INSOLUBLES	---	7940.9	---	1.0	---	0.0	---	0.0	---	211.0	---	0.0	---	8160.9	---	8160.9	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	650.7	4.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	979.0	7.2	979.0	7.2	0.0	0.0
HEMI. XYLAN	132	3857.2	29.2	1.0	.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3858.2	29.2	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	792.3	---	0.0	---	792.3	---	792.3	---	0.0	0.0
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	1453.0	---	0.0	---	1453.0	---	1453.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	85.0	---	85.0	---	85.0	---	0.0	---
EXTRACTIVES	---	691.7	---	1809.4	---	0.0	---	0.0	---	93.0	---	0.0	---	2596.9	---	2596.9	---	0.0	---
SOLUBLE LIGNIN	---	455.3	---	1109.0	---	0.0	---	0.0	---	62.5	---	0.0	---	1706.0	---	1706.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	37.1	0.4	95.7	1.0	0.0	0.0	0.0	0.0	4.9	0.1	0.0	0.0	137.0	1.4	137.0	1.4	0.0	0.0
HMF	126	14.7	0.1	39.1	0.3	0.0	0.0	0.0	0.0	2.0	.0	0.0	0.0	55.7	0.4	55.7	0.4	0.0	0.0
XYLOSE	150	3022.8	20.2	7881.5	52.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	10904.3	72.7	15289.1	101.9	0.0	0.0
GLUCOSE	180	1912.0	10.6	4984.7	27.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6896.6	38.3	43410.9	241.2	0.0	0.0
SULFURIC ACID	98	1021.0	10.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	7.0	0.1	7.0	0.1	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	636.1	15.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	349.0	4.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	200.3	1.4	0.0	0.0	1325.0	9.3	1325.0	9.3	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	51.0	0.6	0.0	0.0	51.0	0.6	51.0	0.6	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	952.6	6.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1317.0	8.3	1317.0	8.3	0.0	0.0
ACETIC ACID	60	276.5	4.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		177109.6		319025.7		983.0		26435.7		104239.1		241.3		628035.1		628035.1		38404.9	
TEMPERATURE, C		90.0		83.0		27.0		27.0		27.0		27.0		50.0		50.0		27.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

DATE 10-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS   SECTION 400: ENZYME HYDROLYSIS

STREAM NO.		410		411		412		413		414		415		416		417		418	
DISKETTE 450 FILE SEC4	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	36573.0	2031.8	1831.9	101.8	524363.7	29131.3	45114.0	2506.3	2258.8	125.5	523936.8	29107.6	15024.3	834.7	62397.1	3466.5	23079.7	1282.2
CELLULOSE	162	0.0	0.0	0.0	0.0	182.7	1.1	3469.3	21.4	173.9	1.1	8.8	0.1	0.0	0.0	3643.2	22.5	3643.2	22.5
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	693.7	---	13170.9	---	659.5	---	34.2	---	1483.1	---	15313.5	---	15313.5	---
ASH	---	0.0	---	0.0	---	8.8	---	158.3	---	8.0	---	0.0	---	18.6	---	185.6	---	185.6	---
MYCELLIUM	---	0.0	---	0.0	---	2.9	---	65.5	---	2.9	---	0.0	---	280.4	---	348.8	---	348.8	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	408.4	---	7752.5	---	387.9	---	20.5	---	872.5	---	9012.8	---	9012.8	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	970.2	7.1	0.0	0.1	0.0	0.0	970.2	7.1	0.0	0.0	8.8	0.1	2.9	.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	785.5	---	6.8	---	0.0	---	785.5	---	2.9	---	9.0	---	1.9	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.9	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	1441.1	---	12.7	---	1.0	---	1440.1	---	12.7	---	26.4	---	9.8	---
CELLOBIASE	---	0.0	---	0.0	---	84.0	---	1.0	---	1.0	---	84.0	---	0.0	---	1.0	---	0.0	---
EXTRACTIVES	---	0.0	---	0.0	---	2574.4	---	22.5	---	0.0	---	2573.4	---	1.0	---	24.4	---	0.8	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	1692.2	---	14.7	---	1.0	---	1691.2	---	0.0	---	15.6	---	5.9	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	136.8	1.4	1.0	.0	0.0	0.0	136.8	1.4	0.0	0.0	1.0	.0	0.0	0.0
HMF	126	0.0	0.0	0.0	0.0	55.7	0.4	0.0	0.0	0.0	0.0	55.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	15155.2	101.0	133.8	0.9	6.8	.0	15148.4	101.0	0.0	0.0	140.7	0.9	51.8	0.3
GLUCOSE	180	0.0	0.0	0.0	0.0	43037.8	239.1	379.1	2.1	19.5	0.1	43020.2	239.0	0.0	0.0	398.6	2.2	147.5	0.8
SULFURIC ACID	98	0.0	0.0	0.0	0.0	7.8	0.1	0.0	0.0	0.0	0.0	7.8	0.1	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	1314.1	9.3	11.7	0.1	1.0	.0	1313.1	9.2	2.0	.0	14.7	0.1	5.9	.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	51.8	0.0	0.0	0.0	0.0	0.0	51.8	0.6	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	1305.3	8.3	11.7	0.1	1.0	.0	1304.3	8.3	0.0	0.0	12.7	0.1	4.9	.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		36573.0		1831.9		594273.9		70334.2		3523.1		592582.7		17697.4		91554.7		51825.0	
TEMPERATURE, C		27.0		27.0		49.0		48.0		47.0		49.0		28.0		44.0		44.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

STREAM NO.		419		420	
DISKETTE 450	MOLECULAR	LBS	MOLES	LBS	MOLES
FILE SEC4	WEIGHTS	HR	HR	HR	HR
WATER	18	39317.4	2184.3	563254.2	31291.9
CELLULOSE	162	0.0	0.0	0.0	0.1
INSOLUBLE LIGNIN	---	0.0	---	34.2	---
ASH	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	20.5	---
YEAST	---	0.0	---	0.0	---
CALCIUM SULFATE	136	5.9	.0	976.0	7.2
HEMI. XYLAN	132	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	5.9	---	791.4	---
NUTRIENTS	---	0.0	---	0.0	---
CELLULASE	---	16.6	---	1456.7	---
CELLOBIASE	---	1.0	---	85.0	---
EXTRACTIVES	---	15.6	---	2589.1	---
SOLUBLE LIGNIN	---	9.8	---	1701.0	---
FUSEL OIL	---	0.0	---	0.0	---
FURFURAL	96	1.0	.0	137.0	1.4
HMF	126	0.0	0.0	55.7	0.4
XYLOSE	150	88.9	0.6	15237.3	101.6
GLUCOSE	180	251.1	1.4	43271.3	240.4
SULFURIC ACID	98	0.0	0.0	7.0	0.1
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	4.8	0.1	1321.9	9.3
SODIUM ACETATE	82	0.0	0.0	51.0	0.6
CALCIUM ACETATE	158	7.8	.0	1312.1	8.3
ACETIC ACID	60	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0
TOTAL		39729.7		632312.4	
TEMPERATURE, C		44.0		48.0	
PRESSURE, ATM		1.0		3.5	

DATE 18-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 500: SUGAR CONCENTRATION

STREAM NO.		501		502		503		504		505		506		507		508		509	
DISKETTE 450 FILE SECS	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	563254.2	31291.9	0.0	0.0	38136.2	2118.7	562803.8	31266.9	514666.0	28592.6	455283.0	25293.5	397052.8	22058.5	336982.0	18721.2	257691.6	14316.2
CELLULOSE	162	8.0	0.1	0.0	0.0	505.1	3.1	513.9	3.2	513.9	3.2	513.9	3.2	513.9	3.2	513.9	3.2	513.9	3.2
INSOLUBLE LIGNIN	---	34.2	---	0.0	---	1963.8	---	1998.0	---	1998.0	---	1998.0	---	1998.0	---	1998.0	---	1998.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	20.5	---	0.0	---	1177.3	---	1197.8	---	1197.8	---	1197.8	---	1197.8	---	1197.8	---	1197.8	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	976.0	7.2	0.0	0.0	29694.9	218.3	30671.0	225.5	30671.0	225.5	30671.0	225.5	30671.0	225.5	30671.0	225.5	30671.0	225.5
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	791.4	---	0.0	---	133.8	---	925.2	---	925.2	---	925.2	---	925.2	---	925.2	---	925.2	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1456.7	---	0.0	---	245.2	---	1701.9	---	1701.9	---	1701.9	---	1701.9	---	1701.9	---	1701.9	---
CELLOBIASE	---	85.0	---	0.0	---	13.7	---	98.7	---	98.7	---	98.7	---	98.7	---	98.7	---	98.7	---
EXTRACTIVES	---	2589.1	---	0.0	---	436.7	---	3025.8	---	3025.8	---	3025.8	---	3025.8	---	3025.8	---	3025.8	---
SOLUBLE LIGNIN	---	1701.0	---	0.0	---	287.2	---	1988.2	---	1988.2	---	1988.2	---	1988.2	---	1988.2	---	1988.2	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	137.8	1.4	0.0	0.0	0.0	0.0	123.1	1.3	104.5	1.1	80.1	0.8	56.7	0.6	32.2	0.3	0.0	0.0
HMF	126	55.7	0.4	0.0	0.0	9.8	0.1	65.5	0.5	65.5	0.5	65.5	0.5	65.5	0.5	65.5	0.5	65.5	0.5
XYLOSE	150	15237.3	101.6	0.0	0.0	2570.5	17.1	17807.8	118.7	17807.8	118.7	17807.8	118.7	17807.8	118.7	17807.8	118.7	17807.8	118.7
GLUCOSE	180	43271.3	240.4	0.0	0.0	7571.8	42.1	50843.1	282.5	50843.1	282.5	50843.1	282.5	50843.1	282.5	50843.1	282.5	50843.1	282.5
SULFURIC ACID	98	7.8	0.1	0.0	0.0	1.0	.0	8.8	0.1	8.8	0.1	8.8	0.1	8.8	0.1	8.8	0.1	8.8	0.1
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	378.1	3.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1321.9	9.3	0.0	0.0	222.8	1.6	1544.6	10.9	1544.6	10.9	1544.6	10.9	1544.6	10.9	1544.6	10.9	1544.6	10.9
SODIUM ACETATE	82	51.8	0.6	0.0	0.0	8.8	0.1	60.6	0.7	60.6	0.7	60.6	0.7	60.6	0.7	60.6	0.7	60.6	0.7
CALCIUM ACETATE	158	1312.1	8.3	0.0	0.0	220.8	1.4	1532.9	9.7	1532.9	9.7	1532.9	9.7	1532.9	9.7	1532.9	9.7	1532.9	9.7
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		632312.4		378.1		83199.4		676910.5		620754.2		569346.0		511093.1		450937.9		371675.2	
TEMPERATURE, C		48.0		35.0		38.0		135.0		127.0		118.0		107.0		88.0		38.0	
PRESSURE, ATM		3.5		1.0		3.1		3.1		2.5		1.9		1.3		0.66		0.065	

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DATE 18-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 500: SUGAR CONCENTRATION

STREAM NO.		510		511		512		513		514		515		516		517		518	
DISKETTE 450 FILE SECS	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	38586.6	2143.7	48137.0	2674.3	59383.0	3299.1	58230.2	3235.0	60070.0	3337.3	79290.4	4405.0	264408.4	14689.4	218091.0	12160.6	664.4	36.9
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	34.2	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	20.5	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	459.2	3.4	516.0	3.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	789.4	---	2.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	1451.8	---	4.9	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	84.0	---	1.0	---
EXTRACTIVES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	2581.2	---	7.6	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	1696.1	---	4.9	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	14.7	0.2	18.6	0.2	24.4	0.3	23.4	0.2	24.4	0.3	32.2	0.3	105.5	1.1	0.0	0.0	0.0	0.0
HMF	126	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	55.7	0.4	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	15193.3	101.3	44.0	0.3
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	43139.4	239.7	131.9	0.7
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	7.0	0.1	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1310.0	9.3	3.9	.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	51.0	0.6	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1308.2	8.3	3.9	.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		38601.3		48156.3		59407.5		58253.6		60095.3		79322.6		264514.0		287027.0		1448.9	
TEMPERATURE, C		135.0		127.0		118.0		107.0		88.0		38.0		88.0		38.0		38.0	
PRESSURE, ATM		3.1		2.5		1.9		1.3		0.66		0.065		0.66		1.0		1.0	

STREAM NO.		519		520		521	
DISKETTE 450 FILE SECS	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	536.4	29.8	128.0	7.1	219427.4	12190.4
CELLULOSE	162	1.0	.0	7.0	.0	1.0	.0
INSOLUBLE LIGNIN	---	1.0	---	33.2	---	1.0	---
ASH	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	1.0	---	19.5	---	1.0	---
YEAST	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	26.4	0.2	490.5	3.6	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	356.6	3.6
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	2.0	---	0.0	---	791.4	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	3.9	---	1.0	---	1455.7	---
CELLOBIASE	---	1.0	---	0.0	---	85.0	---
EXTRACTIVES	---	6.8	---	1.0	---	2588.1	---
SOLUBLE LIGNIN	---	3.9	---	1.0	---	1700.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0
HMF	126	0.0	0.0	0.0	0.0	55.7	0.4
XYLOSE	150	35.2	0.2	8.6	0.1	15228.5	101.5
GLUCOSE	180	106.5	0.6	25.4	0.1	43245.9	240.3
SULFURIC ACID	98	0.0	0.0	0.0	0.0	7.0	0.1
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	2.9	.0	1.0	.0	1828.0	12.9
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	51.8	0.6
CALCIUM ACETATE	158	2.9	.0	1.0	.0	1311.1	8.3
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		730.0		718.1		280135.9	
TEMPERATURE, C		38.0		38.0		38.0	
PRESSURE, ATM		1.0		1.0		1.0	

STREAM NO.	522		523		524	
DISKETTE 450 FILE SECS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	219400.0	12188.9	27.4	1.5	343698.0	19094.4
CELLULOSE	0.0	0.0	1.0	.0	0.0	0.0
INSOLUBLE LIGNIN	0.0	---	1.0	---	0.0	---
ASH	0.0	---	0.0	---	0.0	---
MYCELLIUM	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	0.0	---	1.0	---	0.0	---
YEAST	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	28.3	0.3	328.3	3.3	0.0	0.0
OXYGEN	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	791.4	---	0.0	---	0.0	---
NUTRIENTS	0.0	---	0.0	---	0.0	---
CELLULASE	1455.7	---	0.0	---	0.0	---
CELLOBIASE	85.0	---	0.0	---	0.0	---
EXTRACTIVES	2587.1	---	1.0	---	0.0	---
SOLUBLE LIGNIN	1700.0	---	0.0	---	0.0	---
FUSEL OIL	0.0	---	0.0	---	0.0	---
FURFURAL	0.0	0.0	0.0	0.0	137.6	1.4
HMF	55.7	0.4	0.0	0.0	0.0	0.0
XYLOSE	15226.5	101.5	2.0	.0	0.0	0.0
GLUCOSE	43240.1	240.2	5.9	.0	0.0	0.0
SULFURIC ACID	7.0	0.1	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	1828.0	12.9	0.0	0.0	0.0	0.0
SODIUM ACETATE	51.8	0.6	0.0	0.0	0.0	0.0
CALCIUM ACETATE	1311.1	8.3	0.0	0.0	0.0	0.0
ACETIC ACID	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	287768.5		367.4		343836.6	
TEMPERATURE, C	38.0		38.0		82.0	
PRESSURE, ATM	1.0		1.0		1.0	

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 600: FERMENTATION

STREAM NO.		601		602		603		604	
DISKETTE 450 FILE SEC6	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	219400.0	12188.9	22527.7	1251.5	241927.7	13440.4	0.0	0.0
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	28.3	0.3	0.0	0.0	28.3	0.3	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	791.4	---	0.0	---	791.4	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1455.7	---	0.0	---	1455.7	---	0.0	---
CELLOBIASE	---	85.0	---	0.0	---	85.0	---	0.0	---
EXTRACTIVES	---	2587.1	---	0.0	---	2587.1	---	0.0	---
SOLUBLE LIGNIN	---	1700.0	---	0.0	---	1700.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HMF	126	55.7	0.4	0.0	0.0	55.7	0.4	0.0	0.0
XYLOSE	150	15226.5	101.5	0.0	0.0	15226.5	101.5	0.0	0.0
GLUCOSE	180	43240.1	240.2	0.0	0.0	43240.1	240.2	0.0	0.0
SULFURIC ACID	98	7.0	0.1	0.0	0.0	7.0	0.1	13.7	0.1
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	240.3	5.2	240.3	5.2	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1828.0	12.9	0.0	0.0	1828.0	12.9	0.0	0.0
SODIUM ACETATE	82	51.0	0.6	0.0	0.0	51.0	0.6	0.0	0.0
CALCIUM ACETATE	158	1311.1	8.3	0.0	0.0	1311.1	8.3	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		287760.5		22765.0		310536.5		13.7	
TEMPERATURE, C		30.0		30.0		30.0		27.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0	

STREAM NO.	605		606		607		608		609	
DISKETTE 450 FILE SEC6	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	12168.5	676.0	253606.7	14089.3	489.5	27.2	22070.4	1226.1	32.2	1.8
CELLULOSE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	2396.6	---	2785.4	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	174.9	1.7	203.2	2.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	0.0	0.0	319.5	11.4	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	15.6	0.4	831.4	18.9	20349.9	462.5	0.0	0.0	20349.9	462.5
CORN STEEP LIQUOR	40.1	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	73.3	---	1529.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	4.9	---	89.9	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	130.9	---	2718.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	86.0	---	1786.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	2.0	---	44.9	---	0.0	---	0.0	---	0.0	---
FURFURAL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HMF	2.9	.0	58.6	0.5	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	766.9	5.1	15993.5	106.6	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	1.0	.0	22.5	0.2	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	54.7	0.6	1138.2	12.4	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	1035.6	22.5	21583.9	469.2	245.2	5.3	0.0	0.0	4.9	0.1
SODIUM CARBONATE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	95.7	0.7	1923.7	13.5	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	2.9	0.0	54.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	66.4	0.4	1377.6	8.7	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	26.4	0.6	544.2	12.4	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	17145.4		306610.9		21084.6		22070.4		20387.1	
TEMPERATURE, C	30.0		30.0		30.0		30.0		30.0	
PRESSURE, ATM	1.0		1.0		1.0		1.0		1.0	

EAM NO.		610		611		612		613		614		615		616	
RETIE 450 .E SEC6	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
ATER	18	13435.7	746.4	240171.0	13342.8	12168.5	676.0	1267.2	70.4	182.7	10.1	1084.5	60.2	241255.5	13403.1
LLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
SH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
THER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EAST	---	2645.7	---	139.7	---	2396.6	---	249.1	---	249.1	---	0.0	---	139.7	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	193.4	1.9	9.8	0.1	174.9	1.7	18.6	0.2	18.6	0.2	0.0	0.0	9.8	0.1
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	16.6	0.4	302.9	6.9	15.6	0.4	1.0	.0	0.0	0.0	1.0	.0	303.8	6.9
CORN STEEP LIQUOR	---	44.0	---	787.5	---	40.1	---	3.9	---	1.0	---	2.9	---	790.4	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	81.1	---	1447.9	---	73.3	---	7.8	---	1.0	---	6.0	---	1454.0	---
CELLOBIASE	---	4.9	---	85.0	---	4.9	---	0.0	---	0.0	---	0.0	---	85.0	---
EXTRACTIVES	---	144.6	---	2573.4	---	130.9	---	13.7	---	2.0	---	11.7	---	2585.1	---
SOLUBLE LIGNIN	---	94.0	---	1691.2	---	86.0	---	6.0	---	1.0	---	7.0	---	1699.0	---
FUSEL OIL	---	2.0	---	43.0	---	2.0	---	0.0	---	0.0	---	0.0	---	43.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HMF	126	2.9	.0	55.7	0.4	2.9	.0	0.0	0.0	0.0	0.0	0.0	0.0	55.7	0.4
XYLOSE	150	847.1	5.6	15146.4	101.0	766.9	5.1	80.1	0.5	11.7	0.1	60.4	0.5	15214.8	101.4
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	1.0	.0	21.5	0.2	1.0	.0	0.0	0.0	0.0	0.0	0.0	0.0	21.5	0.2
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	60.6	0.7	1077.6	11.7	54.7	0.6	5.9	0.1	1.0	.0	4.9	0.1	1082.5	11.8
ETHANOL	46	1143.1	24.8	20440.8	444.4	1035.6	22.5	107.5	2.3	15.6	0.3	91.0	2.0	20532.6	446.4
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	105.5	0.7	1818.2	12.8	95.7	0.7	9.6	0.1	1.0	.0	6.0	0.1	1827.0	12.9
SODIUM ACETATE	82	2.9	0.0	51.8	0.0	2.9	0.0	6.8	0.0	1.0	0.0	0.0	0.0	51.8	0.0
CALCIUM ACETATE	158	73.3	0.5	1304.3	8.3	66.4	0.4	0.0	0.0	0.0	0.0	5.9	.0	1310.2	8.3
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	29.3	0.7	514.9	11.7	26.4	0.6	2.9	0.1	0.0	0.0	2.9	0.1	517.8	11.8
TOTAL		18928.4		287682.5		17145.4		1783.0		485.6		1297.5		288980.0	
TEMPERATURE, C		30.0		30.0		30.0		30.0		30.0		30.0		30.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		3.7	

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 700: CARBON DIOXIDE RECOVERY

STREAM NO.		701		702		703	
DISKETTE 450 FILE SEC7	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	32.2	1.8	0.0	0.0	0.0	0.0
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	20349.9	462.5	20349.9	462.5	20349.9	462.5
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0
HMF	126	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	4.9	0.1	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		20387.1		20349.9		20349.9	
TEMPERATURE, C		30.0		30.0		30.0	
PRESSURE, ATM		1.0		1.0		1.0	

DATE 18-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 800: ETHANOL PURIFICATION

STREAM NO.		801		802		803		804		805		806		807		808		809	
DISKETTE 420 FILE SEC8	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	241255.5	13403.1	1457.2	81.0	240873.0	13381.8	7868.0	437.2	1810.4	100.6	6058.4	336.6	849.0	47.2	5209.4	289.4	6058.4	336.6
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	139.7	---	0.0	---	139.7	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	9.8	0.1	0.0	0.0	9.8	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	303.8	6.9	0.0	0.0	0.0	0.0	303.8	6.9	70.3	1.6	233.5	5.3	33.2	0.8	200.3	4.6	233.5	5.3
CORN STEEP LIQUOR	---	790.4	---	0.0	---	790.4	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1454.0	---	0.0	---	1454.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	85.0	---	0.0	---	85.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	2585.1	---	0.0	---	2585.1	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	1699.0	---	0.0	---	1699.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	43.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
HMF	126	55.7	0.4	0.0	0.0	55.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	15214.0	101.4	0.0	0.0	15214.0	101.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	21.5	0.2	0.0	0.0	21.5	0.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	1082.5	11.8	0.0	0.0	1082.5	11.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	20532.6	446.4	20429.9	444.1	102.7	2.2	110321.9	2398.3	25374.6	551.6	84947.2	1846.7	11893.0	258.5	73054.2	1588.1	84947.2	1846.7
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1885.6	13.3	0.0	0.0	1885.6	13.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	51.8	0.0	0.0	0.0	51.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	1310.2	8.3	0.0	0.0	1310.2	8.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	517.6	11.8	0.0	0.0	0.0	0.0	517.6	11.8	119.2	2.7	398.6	9.1	56.7	1.3	342.0	7.8	398.6	9.1
TOTAL		289038.6		21887.1		267361.5		119012.3		27374.6		91637.7		12831.9		78805.8		91637.7	
TEMPERATURE, C		30.0		113.0		138.0		113.0		113.0		110.0		110.0		110.0		110.0	
PRESSURE, ATM		3.7		3.4		3.7		3.4		3.4		3.4		3.4		3.4		3.4	

DATE 10-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS   SECTION 800: ETHANOL PURIFICATION

STREAM NO.		810		811		812		813		814		815		816		817		818	
DISKETTE 420 FILE SEC8	MOLECULAR WEIGHT	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	7868.0	437.2	0.0	0.0	7667.6	426.0	6313.0	350.7	4373.6	243.0	3019.0	167.7	10686.6	593.7	1354.6	75.3	1074.7	59.7
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	-
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	0.0	---	41.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HMF	126	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	76257.6	977.7	76257.6	977.7	616.2	7.9	616.2	7.9	83878.1	1075.4	6.8	0.1	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	110321.9	2398.3	0.0	0.0	20640.4	448.7	20640.4	448.7	3603.9	78.3	3603.9	78.3	24244.3	527.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		118190.6		41.0		104565.6		103211.0		8593.7		7239.1		118809.0		1361.4		1074.7	
TEMPERATURE, C		110.0		27.0		71.0		64.0		64.0		85.0		64.0		43.0		27.0	
PRESSURE, ATM		3.4		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

DATE 16 Aug 84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 800: ETHANOL PURIFICATION

STREAM NO.		819		820		821		822		823		824	
DISKETTE 420 FILE SEC8	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	102.6	5.7	102.6	5.7	0.0	0.0	0.0	0.0	9806.3	544.8	9806.3	544.8
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	303.8	6.9	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	0.0	---	43.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	35.3	0.4	35.3	0.4
HMF	126	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	6.8	0.1	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	20429.9	444.1	20429.9	444.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	11.2	0.2	11.2	0.2
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9.3	0.3	9.3	0.3
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	517.8	11.8	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		20532.5		20575.5		821.7		6.8		9862.1		9862.1	
TEMPERATURE, C		91.0		43.0		30.0		30.0		148.0		148.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		4.4		4.4	

DATE 19-Aug 84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 900: FURFURAL PRODUCTION

STREAM NO.		901		902		903		904		905		906		907		908		909	
DISKETTE 453 FILE SEC9	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	240873.0	13381.8	46131.0	2562.8	0.0	0.0	287004.0	15944.7	292057.4	16225.4	231295.9	12849.8	60761.6	3375.6	221118.5	12284.4	10177.4	565.4
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	4350.6	---	4350.6	---	0.0	---	4350.6	---	0.0	---
YEAST	---	139.7	---	0.0	---	0.0	---	139.7	---	139.7	---	139.7	---	0.0	---	139.7	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	9.8	0.1	0.0	0.0	0.0	0.0	9.8	0.1	9.8	0.1	9.8	0.1	0.0	0.0	9.8	0.1	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	790.4	---	0.0	---	0.0	---	790.4	---	790.4	---	790.4	---	0.0	---	790.4	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1454.8	---	0.0	---	0.0	---	1454.8	---	1454.8	---	1454.8	---	0.0	---	1454.8	---	0.0	---
CELLOBIASE	---	85.0	---	0.0	---	0.0	---	85.0	---	85.0	---	85.0	---	0.0	---	85.0	---	0.0	---
EXTRACTIVES	---	2585.1	---	0.0	---	0.0	---	2585.1	---	2585.1	---	2585.1	---	0.0	---	2585.1	---	0.0	---
SOLUBLE LIGNIN	---	1699.0	---	0.0	---	0.0	---	1699.0	---	1699.0	---	1699.0	---	0.0	---	1699.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	4633.1	48.3	1604.6	16.7	3028.5	31.5	1159.1	12.1	445.5	4.6
HMF	126	55.7	0.4	0.0	0.0	0.0	0.0	55.7	0.4	55.7	0.4	55.7	0.4	0.0	0.0	55.7	0.4	0.0	0.0
XYLOSE	150	15214.8	101.4	0.0	0.0	0.0	0.0	15214.8	101.4	1177.7	7.9	1177.7	7.9	0.0	0.0	1177.7	7.9	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	21.5	0.2	0.0	0.0	3111.7	31.8	3133.2	32.0	3133.2	32.0	3133.2	32.0	0.0	0.0	3133.2	32.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	1082.5	11.8	0.0	0.0	0.0	0.0	1082.5	11.8	1082.5	11.8	1082.5	11.8	0.0	0.0	1082.5	11.8	0.0	0.0
ETHANOL	46	102.7	2.2	0.0	0.0	0.0	0.0	102.7	2.2	102.7	2.2	24.4	0.5	78.3	1.7	0.0	0.0	24.4	0.5
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1827.0	12.9	0.0	0.0	0.0	0.0	1827.0	12.9	1827.0	12.9	1827.0	12.9	0.0	0.0	1827.0	12.9	0.0	0.0
SODIUM ACETATE	82	51.8	0.0	0.0	0.0	0.0	0.0	51.8	0.0	51.8	0.0	51.8	0.0	0.0	0.0	51.8	0.0	0.0	0.0
CALCIUM ACETATE	158	1310.2	8.3	0.0	0.0	0.0	0.0	1310.2	8.3	1310.2	8.3	1310.2	8.3	0.0	0.0	1310.2	8.3	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		267302.9		46131.0		3111.7		316545.7		316545.7		252677.3		63868.3		242030.0		10647.3	
TEMPERATURE, C		138.0		253.0		27.0		220.0		220.0		120.0		120.0		100.0		100.0	
PRESSURE, ATM		3.7		40.8		23.4		23.4		23.4		2.0		2.0		1.0		1.0	

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 900: FURFURAL PRODUCTION

STREAM NO.		910		911		912		913		914		915		916		917		918	
DISKETTE 453 FILE SEC9	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	2366.3	131.5	0.0	0.0	222113.0	12339.6	2523.6	140.2	12701.0	705.6	31983.9	1777.2	190123.1	10562.4	27132.3	1507.3	4857.6	269.9
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	4350.6	---	0.0	---	0.0	---	4350.6	---	0.0	---	217.9	---	4132.7	---
YEAST	---	0.0	---	0.0	---	139.7	---	0.0	---	0.0	---	139.7	---	0.0	---	6.8	---	132.9	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	4347.7	32.0	0.0	0.0	0.0	0.0	4039.9	29.7	307.0	2.3	243.3	1.8	3796.6	27.9
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	9.8	0.1	0.0	0.0	0.0	0.0	9.8	0.1	0.0	0.0	0.0	0.0	9.8	0.1
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	790.4	---	0.0	---	0.0	---	114.3	---	676.1	---	96.7	---	17.6	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	1454.8	---	0.0	---	0.0	---	209.1	---	1245.7	---	177.8	---	31.3	---
CELLOBIASE	---	0.0	---	0.0	---	85.0	---	0.0	---	0.0	---	12.7	---	72.3	---	10.7	---	2.0	---
EXTRACTIVES	---	0.0	---	0.0	---	2585.1	---	0.0	---	0.0	---	372.2	---	2212.9	---	315.6	---	56.7	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	1699.0	---	0.0	---	0.0	---	244.3	---	1454.8	---	207.1	---	37.1	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	1090.7	11.4	68.4	0.7	513.9	5.4	157.3	1.6	933.4	9.7	133.8	1.4	23.4	0.2
HMF	126	0.0	0.0	0.0	0.0	55.7	0.4	0.0	0.0	0.0	0.0	7.8	0.1	47.9	0.4	6.8	0.1	1.0	.0
XYLOSE	150	0.0	0.0	0.0	0.0	1177.7	7.9	0.0	0.0	0.0	0.0	170.0	1.1	1007.7	6.7	144.6	1.0	25.4	0.2
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	2366.3	32.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	1082.5	11.8	0.0	0.0	0.0	0.0	156.3	1.7	926.2	10.1	132.9	1.4	23.4	0.3
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	24.4	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	1827.0	12.9	0.0	0.0	0.0	0.0	271.6	1.9	1555.4	11.0	230.6	1.6	41.0	0.3
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	51.0	0.0	0.0	0.0	0.0	0.0	7.8	0.0	44.0	0.0	6.8	0.0	1.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	1310.2	8.3	0.0	0.0	0.0	0.0	188.6	1.2	1121.6	7.1	160.2	1.0	28.3	0.2
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		2366.3		2366.3		244170.6		2592.0		13239.3		42441.9		201728.7		29224.0		13217.8	
TEMPERATURE, C		27.0		27.0		100.0		100.0		100.0		100.0		100.0		100.0		100.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0		1.0	

DATE 19-Aug-86

ETHANOL VIA ENZYME HYDROLYSIS   SECTION 900: FURFURAL PRODUCTION

STREAM NO.		919		920		921		922		923		924		925		926		927	
DISKETTE 453 FILE SEC9	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	217255.4	12069.7	73462.6	4081.3	62399.0	3466.6	290616.9	16145.4	62297.9	3461.0	957.9	53.2	935.0	51.9	62376.2	3465.3	78.3	4.3
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	217.9	---	0.0	---	0.0	---	217.9	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	6.8	---	0.0	---	0.0	---	6.8	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	551.0	4.1	0.0	0.0	0.0	0.0	551.0	4.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	772.8	---	0.0	---	0.0	---	772.8	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1423.5	---	0.0	---	0.0	---	1423.5	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	83.0	---	0.0	---	0.0	---	83.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	2520.5	---	0.0	---	0.0	---	2520.5	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	1661.9	---	0.0	---	0.0	---	1661.9	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	1067.3	11.1	3542.4	36.9	18614.0	193.9	46.1	0.5	14051.2	146.4	5042.3	52.5	503.2	5.2	14075.6	146.6	24.4	0.3
HMF	126	56.7	0.4	0.0	0.0	0.0	0.0	56.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	1152.3	7.7	0.0	0.0	0.0	0.0	1152.3	7.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	1059.1	11.5	0.0	0.0	0.0	0.0	1059.1	11.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	102.7	2.2	102.7	2.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	102.7	2.2	102.7	2.2
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1786.0	12.6	0.0	0.0	0.0	0.0	1786.0	12.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	50.0	0.6	0.0	0.0	0.0	0.0	50.0	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	1281.0	8.1	0.0	0.0	0.0	0.0	1281.0	8.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		230954.7		77107.7		81116.5		303295.0		76349.1		6000.1		1438.1		76554.5		205.4	
TEMPERATURE, C		100.0		100.0		98.0		104.0		98.0		98.0		98.0		98.0		80.0	
PRESSURE, ATM		1.0		1.0		1.0		3.1		1.0		1.0		1.0		1.0		1.0	

STREAM NO.		920	
DISKETTE 453 FILE SEC9	MOLECULAR WEIGHTS	LBS HR	MOLES HR
WATER	18	22.9	1.3
CELLULOSE	162	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---
ASH	---	0.0	---
MYCELLIUM	---	0.0	---
OTHER INSOLUBLES	---	0.0	---
YEAST	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0
HEMI. XYLAN	132	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0
OXYGEN	32	0.0	0.0
NITROGEN	28	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---
NUTRIENTS	---	0.0	---
CELLULASE	---	0.0	---
CELLOBIASE	---	0.0	---
EXTRACTIVES	---	0.0	---
SOLUBLE LIGNIN	---	0.0	---
FUSEL OIL	---	0.0	---
FURFURAL	96	4539.1	47.3
HMF	126	0.0	0.0
XYLOSE	150	0.0	0.0
GLUCOSE	180	0.0	0.0
SULFURIC ACID	98	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0
BENZENE	78	0.0	0.0
GLYCEROL	92	0.0	0.0
ETHANOL	46	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0
SODIUM SULFATE	142	0.0	0.0
SODIUM ACETATE	82	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0
ACETIC ACID	60	0.0	0.0
METHANOL	32	0.0	0.0
ACETALDEHYDE	44	0.0	0.0
TOTAL		4562.0	
TEMPERATURE, C		43.0	
PRESSURE, ATM		1.0	

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ETHANOL VIA ENZYME HYDROLYSIS SECTION 1000: HEAT GENERATION

STREAM NO.		1001		1002		1003		1004		1005		1006		1007		1008		1009	
DISKETTE 453 FILE SEC10	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	290616.9	16145.4	23079.7	1282.2	182.7	10.1	78.3	4.3	48914.3	2717.5	257008.1	14278.2	218709.7	12150.5	176796.4	9822.0	131170.5	7287.2
CELLULOSE	162	0.0	0.0	3643.2	22.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	15313.5	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	185.6	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	348.8	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	217.9	---	9012.0	---	0.0	---	0.0	---	0.0	---	217.9	---	217.9	---	217.9	---	217.9	---
YEAST	---	6.0	---	0.0	---	249.1	---	0.0	---	0.0	---	6.0	---	6.0	---	6.0	---	6.0	---
CALCIUM SULFATE	136	551.0	4.1	2.9	.0	0.0	0.0	0.0	0.0	0.0	0.0	551.0	4.1	551.0	4.1	551.0	4.1	551.0	4.1
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	10.6	0.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	772.0	---	3.9	---	1.0	---	0.0	---	0.0	---	772.0	---	772.0	---	772.0	---	772.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1423.5	---	9.8	---	1.0	---	0.0	---	0.0	---	1423.5	---	1423.5	---	1423.5	---	1423.5	---
CELLOBIASE	---	81.0	---	0.0	---	0.0	---	0.0	---	0.0	---	83.0	---	81.0	---	81.0	---	81.0	---
EXTRACTIVES	---	2520.5	---	0.0	---	2.0	---	0.0	---	0.0	---	2520.5	---	2520.5	---	2520.5	---	2520.5	---
SOLUBLE LIGNIN	---	1661.9	---	5.9	---	1.0	---	0.0	---	0.0	---	1661.9	---	1661.9	---	1661.9	---	1661.9	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	46.1	0.5	0.0	0.0	0.0	0.0	24.4	0.3	175.7	1.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HMF	126	56.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	56.7	0.4	56.7	0.4	56.7	0.4	56.7	0.4
XYLOSE	150	1152.3	7.7	51.0	0.3	11.7	0.1	0.0	0.0	0.0	0.0	1152.3	7.7	1152.3	7.7	1152.3	7.7	1152.3	7.7
GLUCOSE	180	0.0	0.0	147.5	0.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	1059.1	11.5	0.0	0.0	1.0	.0	0.0	0.0	0.0	0.0	1059.1	11.5	1059.1	11.5	1059.1	11.5	1059.1	11.5
ETHANOL	46	0.0	0.0	0.0	0.0	15.6	0.3	102.7	2.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1786.0	12.6	5.9	.0	1.0	.0	0.0	0.0	0.0	0.0	1786.0	12.6	1786.0	12.6	1786.0	12.6	1786.0	12.6
SODIUM ACETATE	82	50.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	0.0	0.0	50.0	0.0	50.0	0.0	50.0	0.0	50.0	0.0
CALCIUM ACETATE	158	1281.0	8.1	4.9	.0	0.0	0.0	0.0	0.0	0.0	0.0	1281.0	8.1	1281.0	8.1	1281.0	8.1	1281.0	8.1
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	56.2	0.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	46.4	1.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		303295.0		51825.0		485.6		205.4		49192.5		269640.1		231341.7		189420.4		143802.5	
TEMPERATURE, C		164.0		44.0		30.0		80.0		148.0		135.0		127.0		118.0		107.0	
PRESSURE, ATM		3.1		1.0		1.0		1.0		4.4		3.1		2.5		1.9		1.3	

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 1000: HEAT GENERATION

STREAM NO.		1010		1011		1012		1013		1014		1015		1016		1017		1018	
DISKETTE 453 FILE SEC10	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	76653.9	4258.5	12631.8	701.8	33608.8	1867.2	38298.4	2127.7	41913.3	2328.5	45625.9	2534.8	54516.6	3028.7	64022.0	3556.8	326899.3	18161.1
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	217.9	---	217.9	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
YEAST	---	6.8	---	6.8	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CALCIUM SULFATE	136	551.0	4.1	551.0	4.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	772.8	---	772.8	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	1423.5	---	1423.5	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLOBIASE	---	83.0	---	83.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
EXTRACTIVES	---	2528.5	---	2528.5	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
SOLUBLE LIGNIN	---	1661.9	---	1661.9	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	46.1	0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	221.5	2.3
HMF	126	56.7	0.4	56.7	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
XYLOSE	150	1152.3	7.7	1152.3	7.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	1059.1	11.5	1059.1	11.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1786.0	12.6	1786.0	12.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM ACETATE	82	50.8	0.0	50.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM ACETATE	158	1281.8	8.1	1281.8	8.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	56.2	0.9
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	46.4	1.5
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		89285.9		25263.9		33654.9		38298.4		41913.3		45625.9		54516.6		64022.0		327223.7	
TEMPERATURE, C		88.0		38.0		135.0		127.0		118.0		103.0		88.0		38.0		76.0	
PRESSURE, ATM		0.66		0.065		3.1		2.5		1.9		1.3		0.66		0.065		1.0	

STREAM NO.		1019		1020		1021		1022	
DISKETTE 453 FILE SEC10	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	11577.5	643.2	0.0	0.0	47550.0	2641.7	0.0	0.0
CELLULOSE	162	5464.6	33.7	0.0	0.0	9107.0	56.2	0.0	0.0
INSOLUBLE LIGNIN	---	2130.2	---	0.0	---	17443.7	---	0.0	---
ASH	---	23.1	---	0.0	---	200.0	---	200.0	---
MYCELLIUM	---	0.0	---	0.0	---	340.0	---	0.0	---
OTHER INSOLUBLES	---	1227.3	---	0.0	---	10457.9	---	500.2	---
YEAST	---	0.0	---	0.0	---	256.0	---	0.0	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	554.0	4.1	554.0	4.1
HEMI. XYLAN	132	1875.6	14.2	0.0	0.0	1875.6	14.2	0.0	0.0
HEMI. GLUCAN	162	532.5	3.3	0.0	0.0	532.5	3.3	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	18.6	0.2	0.0	0.0
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	303.0	6.9	303.0	6.9	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	777.7	---	0.0	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	1434.2	---	0.0	---
CELLOBIASE	---	0.0	---	0.0	---	83.0	---	0.0	---
EXTRACTIVES	---	324.2	---	0.0	---	2863.4	---	0.0	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	1668.7	---	0.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	0.0	0.0	24.4	0.3	0.0	0.0
HMF	126	0.0	0.0	0.0	0.0	56.7	0.4	0.0	0.0
XYLOSE	150	0.0	0.0	0.0	0.0	1215.0	8.1	0.0	0.0
GLUCOSE	180	0.0	0.0	0.0	0.0	147.5	0.8	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	1060.0	11.5	0.0	0.0
ETHANOL	46	0.0	0.0	0.0	0.0	118.3	2.6	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	1792.0	12.6	1792.0	12.6
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	51.0	0.6	0.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	1286.7	8.1	0.0	0.0
ACETIC ACID	60	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	517.0	11.8	517.0	11.8	0.0	0.0
TOTAL		23155.0		821.7		101756.4		3055.8	
TEMPERATURE, C		30.0		30.0		30.0		80.0	
PRESSURE, ATM		1.0		1.0		1.0		1.0	

DATE 19-Aug-84

ETHANOL VIA ENZYME HYDROLYSIS SECTION 1100: WASTE TREATMENT

STREAM NO.		1101		1102		1103		1104		1105		1106		1107		1108		1109	
DISKETTE 453 FILE SECT1	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	44387.1	2465.9	343698.8	19094.4	9806.3	544.8	1354.6	75.3	326899.3	18161.1	616.5	34.2	128.0	7.1	27.4	1.5	4857.6	269.9
CELLULOSE	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	99.7	0.6	7.8	.0	1.0	.0	0.0	0.0
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	37.1	---	33.2	---	1.0	---	0.0	---
ASH	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
MYCELLIUM	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	21.5	---	19.5	---	1.0	---	4132.7	---
YEAST	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	132.9	---
CALCIUM SULFATE	136	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3079.5	22.6	490.5	3.6	0.0	0.0	3796.6	27.9
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9.8	0.1	0.0	0.0	0.0	0.0	0.0	0.0
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	328.3	3.3	9.8	0.1
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	17.6	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	1.0	---	0.0	---	31.3	---
CELLOBIASE	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	2.0	---
EXTRACTIVES	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	3.9	---	1.0	---	1.0	---	56.7	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	2.9	---	1.0	---	0.0	---	37.1	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	135.9	1.4	137.8	1.4	35.3	0.4	0.0	0.0	221.8	2.3	0.0	0.0	0.0	0.0	0.0	0.0	23.4	0.2
HMF	126	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	.0
XYLOSE	150	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	16.6	0.1	8.8	0.1	2.0	.0	25.4	0.2
GLUCOSE	180	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	10.7	0.1	25.4	0.1	5.9	.0	0.0	0.0
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	0.0	0.0	0.0	0.0	6.8	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
GLYCEROL	92	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	23.4	0.3
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	.0	0.0	0.0	41.0	0.3
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.0	.0	1.0	.0	0.0	0.0	28.3	0.2
ACETIC ACID	60	54.7	0.9	0.0	0.0	11.2	0.2	0.0	0.0	56.2	0.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
METHANOL	32	11.7	0.4	0.0	0.0	9.3	0.3	0.0	0.0	46.4	1.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		44587.4		343836.6		9862.1		1361.4		327223.7		3900.2		718.1		367.4		13217.8	
TEMPERATURE, C		100.0		82.0		140.0		43.0		76.0		81.0		38.0		38.0		100.0	
PRESSURE, ATM		1.1		1.0		4.4		1.0		1.0		1.0		1.0		1.0		1.0	

STREAM NO.		1110		1111		1112	
DISKETTE 453 FILE SEC11	MOLECULAR WEIGHTS	LBS HR	MOLES HR	LBS HR	MOLES HR	LBS HR	MOLES HR
WATER	18	0.0	0.0	726146.2	40341.5	5629.5	312.7
CELLULOSE	162	0.0	0.0	0.0	0.0	108.4	0.7
INSOLUBLE LIGNIN	---	0.0	---	0.0	---	71.3	---
ASH	---	208.8	---	0.0	---	0.0	---
MYCELIUM	---	0.0	---	0.0	---	0.0	---
OTHER INSOLUBLES	---	500.2	---	0.0	---	4174.7	---
YEAST	---	0.0	---	0.0	---	132.9	---
CALCIUM SULFATE	136	554.0	4.1	0.0	0.0	7366.6	54.2
HEMI. XYLAN	132	0.0	0.0	0.0	0.0	9.0	0.1
HEMI. GLUCAN	162	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM CARBONATE	100	0.0	0.0	0.0	0.0	338.0	3.4
OXYGEN	32	0.0	0.0	0.0	0.0	0.0	0.0
NITROGEN	28	0.0	0.0	0.0	0.0	0.0	0.0
CARBON DIOXIDE	44	0.0	0.0	0.0	0.0	0.0	0.0
CORN STEEP LIQUOR	---	0.0	---	0.0	---	17.6	---
NUTRIENTS	---	0.0	---	0.0	---	0.0	---
CELLULASE	---	0.0	---	0.0	---	32.2	---
CELLOBIASE	---	0.0	---	0.0	---	2.0	---
EXTRACTIVES	---	0.0	---	0.0	---	62.5	---
SOLUBLE LIGNIN	---	0.0	---	0.0	---	41.0	---
FUSEL OIL	---	0.0	---	0.0	---	0.0	---
FURFURAL	96	0.0	0.0	530.7	5.5	23.4	0.2
HMF	126	0.0	0.0	0.0	0.0	1.0	.0
XYLOSE	150	0.0	0.0	0.0	0.0	52.0	0.4
GLUCOSE	180	0.0	0.0	0.0	0.0	42.0	0.2
SULFURIC ACID	98	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM HYDROXIDE	40	0.0	0.0	0.0	0.0	0.0	0.0
CALCIUM HYDROXIDE	74	0.0	0.0	0.0	0.0	0.0	0.0
BENZENE	78	0.0	0.0	6.0	0.1	6.0	0.1
GLYCEROL	92	0.0	0.0	0.0	0.0	23.4	0.3
ETHANOL	46	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM CARBONATE	106	0.0	0.0	0.0	0.0	0.0	0.0
SODIUM SULFATE	142	1792.0	12.6	0.0	0.0	42.0	0.3
SODIUM ACETATE	82	0.0	0.0	0.0	0.0	1.0	0.0
CALCIUM ACETATE	158	0.0	0.0	0.0	0.0	31.3	0.2
ACETIC ACID	60	0.0	0.0	122.1	2.0	0.0	0.0
METHANOL	32	0.0	0.0	67.4	2.1	0.0	0.0
ACETALDEHYDE	44	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL		3055.0		726873.2		18210.3	
TEMPERATURE, C		80.0		81.0			
PRESSURE, ATM		1.0		1.0			

B. Major Equipment Specifications and Costs

This sub-section contains the major equipment specifications for the eleven plant sections based on a capacity of 25 million gallons per year of fuel grade ethanol with an on-stream factor of 8,000 hours per year. The equipment sparing philosophy was designed to achieve this on-stream factor. Bare equipment costs were developed in part from vendor quotations and in part from an Icarus Cost computer run.

EQUIPMENT SPECIFICATIONS AND COSTS

SECTION 100 - PRETREATMENT/PREHYDROLYSIS

	Cost
Conveyor (CO-101)	
Part of offsite wood handling	-
Chip Washer (VS-101) - 3 Items	
Part of offsite wood handling	-
Conveyor (CO-102) -3 Items	
Part of offsite wood handling	-
Chip Bin (VT-101) - 3 Items	$69,300
Type: Cone bottom Capacity: 1,600 cubic feet Size: 9' dia. x 23' t/t Material: CS Temperature: 100^{o}F Pressure: 0.05 psig	
Feeder (CO-103) - 3 Items	
Part of refiner (CR-101) package	-
Digester (CO-104) - 3 Items	
Part of refiner (CR-101) package	-

Disk Refiner (CR-101) - 3 Items — $1,215,000

Type: Single disk
Capacity: 380 dry tons per day
Drive: 3,000 HP
Temperature: 392°F
Pressure: 220 psig

Prehydrolysis Reactor (R-101) - 3 Items — $55,300

Capacity: 80 gallons
Size: 1' dia. x 13.5' t/t
Material: Zirconium clad
Internal: Discharge wiper
Drive: 1.5 HP
Temperature: 392°F
Pressure: 210 psig

First Stage Water Preheater (HE-101) — $12,200

Area; 840 square feet
Type: Fixed T-S
Duty: 17.3 MM Btu/hr
Material: CS/CS
Temperature: 250/150°F
Pressure: 20/150 psig

Second Stage Water Preheater (HE-102) — $26,100

Area: 2,100 square feet
Type: Fixed T-S
Duty: 26.2 MM Btu/hr
Material: CS/CS
Temperature: 250/150°F
Pressure: 30/150 psig

Flash Vessel (VT-102) — $124,600

Type: Vertical
Capacity: 6,100 gallons
Size: 7' dia. x 21' t/t
Material: Zirconium
Internal: Demister
Temperature: 300°F
Pressure: 50 psig

Flash Vessel (VT-103) — $93,700

Type: Vertical
Capacity: 3,800 gallons
Size: 6' dia. x 18' t/t
Material: Zirconium
Temperature: 250°F
Pressure: 15 psig

Flash Vessel (VT-104)	$40,400

Type: Vertical
Capacity: 12,900 gallons
Size: 9' dia. x 27' t/t
Material: 317 SS
Temperature: 212^{o}F
Pressure: 2 psig

Water Feed Pump (CP-101) - 2 Items	$15,800

Type: Centrifugal, AVS Medium
Capacity: 750 GPM
SG: 1.0
Material: CS
Head: 575' TDH
Drive: 150 HP
Temperature: 80^{o}F

Subtotal Section 100	$1,652,400
Undefined miscellaneous	30,600
Total Section 100	$1,683,000

SECTION 200 - SUGAR SEPARATION AND NEUTRALIZATION

	Cost
Centrifuge (CT-201) - 2 Items	$911,000

Type: Solid bowl
Size: 44" dia.
Material: 317 SS
Drive: 25 HP
Overflow: 311 GPM

Prehydrolyzate Bin (VT-201)	$143,200

Type: Cone bottom
Capacity: 90,000 gallons
Material: 317 SS
Temperature: 212°F
Pressure: 0.05 psig
Holdup: 4 hours

Overflow Pump (CP-201) - 4 Items	$37,600

Type: Centrifugal, AVS medium
Capacity: 310 GPM
Material: 317 SS
Head: 100' TDH
Drive: 15 HP
Temperature: 212°F

Rotary Drum Polishing Filter (F-201)	$50,200

Area: 100 square feet
Capacity: 622 GPM
Material: EPLCS
Drive: 3 HP

Thickener Exchanger (HE-201)	$55,300

Area: 1,175 square feet
Type: Fixed T-S
Duty: 9.4 MM Btu/hr
Material: 316 SS/321 S
Temperature: 185/212°F
Pressure: 15/15 psig

Filter Cooler (HE-202)	$34,100

Area: 880 square feet
Type: Fixed T-S
Duty: 12.4 MM Btu/hr
Material: CS/321 S
Temperature: 120/185°F
Pressure: 15/15 psig

Lime Tank (AOT-201) $87,000

Type: Mixer
Capacity: 11,000 gallons
Material: 304 SS
Drive: 50 HP
Holdup: 4 hours

Neutralization Tank (AOT-202) $66,100

Type: Mixer
Capacity: 7,000 gallons
Material: 317 SS
Holdup: 10 minutes
Drive: 40 HP

Lime Pump (CP-202) - 2 Items $5,400

Type: Centrifugal, AVS medium
Capacity: 50 GPM
Material: 304 SS
Head: 100' TDH
Drive: 2 HP
Temperature: 80°F

Thickener Feed Pump (CP-203) - 2 Items $8,800

Type: Centrifugal, AVS medium
Capacity: 700 GPM
Material: 304 SS
Head: 100' TDH
Drive: 25 HP
Temperature: 140°F

Conveyor (CO-201) $174,100

Type: Screw
Material: 317 SS
Drive: 40 HP
Length: 100'
Screw Diameter: 24"

Conveyor (CO-202) $50,700

Type: Screw
Material: 317 SS
Drive: 7.5 HP
Length: 100'
Screw Diameter: 9"

Thickener (T-201)	$41,000

Type: Gravity w/rake
Capacity: 650 GPM
Material: CS
Drive: 10 HP
Diameter: 30'

Centrifuge (CT-202)	$71,600

Type: Solid bowl
Size: 18" dia.
Material: 304 SS
Drive: 25 HP
Overflow: 27 GPM

Overflow Pump (CP-204) - 2 Items	$5,400

Type: Centrifugal, AVS medium
Capacity: 30 GPM
Material: 304 SS
Head: 100' TDH
Drive: 3 HP
Temperature: 180°F

Sugar Receiver (VT-202)	$167,200

Type: Cone roof
Capacity: 325,000 gallons
Material: 304 SS
Temperature: 180°F
Pressure: 0.05 psig
Holdup: 8 hours

Sugar Transfer Pump (CP-205) - Items	$8,300

Type: Centrifugal, AVS medium
Capacity: 725 GPM
Material: 304 SS
Head: 100' TDH
Drive: 25 HP
Temperature: 180°F

Subtotal Section 200	$1,918,000
Undefined miscellaneous	134,400
Total Section 200	$2,052,400

SECTION 300 - ENZYME PRODUCTION

	Cost
Solids Bin (VT-318)	$13,500

Type: Cone bottom
Capacity: 12,000 gallons
Material: 317 SS
Temperature: 210°F
Pressure: 0.05 psig
Holdup: 4 hours

Conveyor (CO-301)	$50,700

Type: Screw
Capacity: 13 tons/hour
Material: 316 SS
Drive: 7.5 HP
Length: 100'
Screw Diameter: 9"

Enzyme Production Tanks (VT-301-314) - 14 Items	$1,695,400

Type: Cone roof
Capacity: 185,000 gallons
Material: 304 SS
Temperature: 85°F
Pressure: 0.05 psig

Coolers (HE-301-314) - 14 Items	$71,400

Area: 40 square feet
Type: Spiral plate
Duty: 0.225 MM Btu/hr
Material: 304 SS
Pressure: 15 psig

Recirculation Pumps (CP-301-314) - 21 Items	$98,700

Type: Centrifugal, recessed impeller
Capacity: 50 GPM
Material: 304 SS
Head: 100' TDH
Drive: 20 HP
Temperature: 85°F

Filter (F-301)	$2,000

Area: 50 square feet
Type: Cartridge
Material: CS

Filter (F-302) $2,000

Area: 50 square feet
Type: Cartridge
Material: CS

Air Blower (FN-301) $10,000

Type: ROT-Blower
Capacity: 100 GPM
Material: CS
Drive: 10 HP

Discharge Pumps (CP-315-328) - 21 Items 98,700

Type: Centrifugal, recessed impeller
Capacity: 50 GPM
Material: 304 SS
Head: 100' TDH
Drive: 20 HP
Temperature: 85°F

Hold Tank (HT-301) $152,600

Type: Horizontal drum
Capacity: 290,000 gallons
Material: 304 SS
Temperature: 85°F
Pressure: 0.05 psig
Holdup: 36 hours

Cell Centrifuge Feed Pump (CP-329) - 2 Items 9,400

Type: Centrifugal, recessed impeller
Capacity: 150 GPM
Material: 304 SS
Head: 100' TDH
Drive: 20 HP
Temperature: 35°F

Cell Centrifuge (CT-301) 56,400

Type: Disk
Size: 10" diameter
Material: CS
Drive: 25 HP
Overflow: 50 GPM

Repulping Tank (AOT-301) $19,500

Type: Mixer
Capacity: 1,500 gallons
Material: 304 SS-clad

Drive: 5 HP
Temperature: 85°F
Pressure: 0.05 psig
Holdup: 10 minutes

<u>Wash Water Tank (VT-315)</u> $17,600

Type: Cylinder
Capacity: 10,000 gallons
Material: CS
Temperature: 80°F
Pressure: 0.05 psig
Holdup: 1 hour

<u>Wash Water Pump (CP-330) - 2 Items</u> $7,100

Type: Centrifugal, AVS medium
Capacity: 160 GPM
Material: CS
Head: 100' TDH
Drive: 10 HP
Temperature: 80°F

<u>Polishing Centrifuge (CT-302)</u> $143,200

Type: Disk
Size: 20" dia.
Material: CS
Drive: 150 HP
Overflow: 160 GPM

<u>Centrifuge Feed Pump (CP-331) - 2 Items</u> $6,800

Type: Centrifugal, AVS medium
Capacity: 125 GPM
Material: 304 SS
Head: 100' TDH
Drive: 5 HP
Temperature: 85°F

<u>Cell Recycle Hold Tank (VT-317)</u> $115,800

Type: Cone roof
Capacity: 175,000 gallons
Material: 304 SS
Temperature: 85°F
Pressure: 0.05 psig
Holdup: 36 hours

<u>Cell Recycle Pump (CP-332) - 2 Items</u> $6,200

Type: Centrifugal, AVS medium
Capacity: 90 GPM

Material: 304 SS
Head: 100' TDH
Drive: 5 HP
Temperature: 85°F

Enzyme Receiver (VT-316)	$192,500

Type: Cone roof
Capacity: 410,000 gallons
Material: 304 SS
Temperature: 35°F
Pressure: 0.05 psig
Holdup: 36 hours

Enzyme Feed Pump (CP-333) - 2 Items	$7,800

Type: Centrifugal, AVS medium
Capacity: 220 GPM
Material: 304 SS
Head: 100' TDH
Drive: 7.5 HP
Temperature: 85°F

Subtotal Section 300	$2,777,300
Undefined Miscellaneous	194,700
Total Section 300	$2,972,000

SECTION 400 - ENZYME HYDROLYSIS

	Cost
Slurry Tank (AOT-401)	$99,400

Type: Mixer
Capacity: 15,000 gallons
Material: 317 SS
Drive: 50 HP
Temperature: 190°F
Pressure: 0.05 psig
Holdup: 10 minutes

Slurry Cooler (HE-407)	$305,400

Area: 11,000 square feet
Type: Fixed T-S
Duty: 24.4 MM Btu/hr
Material: CS/304 SS
Temperature: 115/165°F
Pressure: 15/15 psig

Feed Pump (CP-413) - 2 Items	$13,800

Type: Centrifugal, recessed impeller
Capacity: 1,250 GPM
Material: 304 SS
Head: 100' TDH
Drive: 100 HP

Enzyme Hydrolysis Tanks (VT-401-406) - 6 Items	$1,077,600

Type: Cone roof
Capacity: 370,000 gallons
Material: 304 SS
Temperature: 122°F
Pressure: 0.05 psig

Agitators (A-401-406) - 6 Items	$120,000

Type: Propeller
Capacity: 40,000 GPM
Material: 304 SS
Size: 30"
Drive: 75 HP
Temperature: 122°F

Discharge Pumps (CP-407-412) - 9 Items	$62,100

Type: Centrifugal, recessed impeller
Capacity: 1,250 GPM

Material: 304 SS
Head: 100' TDH
Drive: 100 HP
Temperature: 122°F

Steam Coils (HE-401-406) - 6 Items	$25,200

Area: 100 square feet
Type: Heater steam
Material: 304 SS
Pressure: 15 psig

Centrifuge (CT-401) - 3 Items	$1,239,600

Type: Solid bowl
Size: 54" dia.
Material: 304 SS
Drive: 250 HP
Overflow: 400 GPM

Overflow Pump (CP-415) - 6 Items	$23,400

Type: Centrifugal, AVS medium
Capacity: 450 GPM
Material: 304 SS
Head: 100' TDH
Drive: 20 HP
Temperature: 120°F

Rotary Drum Polishing Filter (F-401)	$40,200

Area: 100 square feet
Type: Rotary drum
Capacity: 1,200 GPM
Material: 304 SS
Drive: 3 HP

Wash Water Tank (VT-407)	$12,000

Type: Cylinder
Capacity: 5,000 gallons
Material: CS
Temperature: 80°F
Pressure: 0.05 psig

Wash Water Pump (CP-414) - 2 Items	$6,000

Type: Centrifugal, AVS medium
Capacity: 80 GPM
Material: CS
Head: 100' TDH
Drive: 5 HP
Temperature: 80°F

Conveyor (CO-401)	$12,400
Type: Screw Capacity: 30 tons/hour Material: CS Drive: 15 HP Length: 100' Screw diameter: 12"	
Dewatering Press (F-402) - 3 Items	$618,000
Type: Fibercone press Capacity: 60 GPM Material: 304 SS Drive: 75 HP	
Press Discharge Pump (CP-415) - 2 Items	$6,400
Type: Centrifugal, AVS medium Capacity: 100 GPM Material: 304 SS Head: 100' TDH Drive: 5 HP Temperature: 115°F	
Sugar Hold Tank (VT-408)	$292,000
Type: Cone roof Capacity: 630,000 gallons Material: 304 SS Temperature: 120°F Pressure: 0.05 psig Holdup: 8 hours	
Sugar Transfer Pump (CP-417) - 2 Items	$13,800
Type: Centrifugal, AVS medium Capacity: 1,300 GPM Material: 304 SS Head: 150' TDH Drive: 50 HP Temperature: 120°F	
Subtotal Section 400	$3,967,300
Undefined Miscellaneous	277,700
Total Section 400	$4,245,000

SECTION 500 - SUGAR CONCENTRATION

	Cost
Multi-Effect-Evaporator (E-501)	$3,260,000

Type: Six effect-forced circulation
Duty: Heating - 68.6 MM Btu/hr
Cooling - 82.3 MM Btu/hr
Capacity: 344,000 lb/hr distillate
Material: 304 SS
Drive: 1,200 HP

$CaSO_4$ Centrifuge (CT-501)	$10,000

Type: Solid bowl
Size: 5" diameter
Material: 304 SS
Drive: 1 HP
Overflow: 2 GPM

Sugar Pump (CP-501) - 2 Items	$8,200

Type: Centrifuge, AVS medium
Capacity: 550 GPM
Material: 304 SS
Head: 100' TDH
Drive: 20 HP
Temperature: 100°F

Overflow Pump (CP-502) - 2 Items	$4,000

Type: Centrifugal, AVS medium
Capacity: 2 GPM
Material: 304 SS
Head: 50' TDH
Drive: 0.5 HP
Temperature: 100°F

Precipitator (AOT-501)	$45,300

Type: Mixer
Capacity: 6,500 gallons
Material: 304 SS
Drive: 40 HP
Temperature: 100°F
Pressure: 0.05 psig
Holdup: 10 minutes

Na_2CO_3 Solution Tank (AOT-502)	$7,000

Type: Mixer
Capacity: 50 gallons

Material: 304 SS
Drive: 0.5 HP
Temperature: 100°F
Pressure: 0.05 psig

Na_2CO_3 Feed Pump (CP-503) - 2 Items	$4,000

Type: Centrifugal, AVS medium
Capacity: 5 GPM
Material: 304 SS
Head: 20' TDH
Drive: 0.5 HP
Temperature: 100°F

$CaCO_3$ Centrifuge (CT-502)	$336,000

Type: Solid bowl
Size: 44" dia.
Material: 304 SS
Drive: 250 hp
Overflow: 642 GPM

Overflow Pump (CP-504) - 2 Items	$8,400

Type: Centrifugal, AVS medium
Capacity: 600 GPM
Material: 304 SS
Head: 100' TDH
Drive: 25 HP
Temperature: 100°F

Filter Feed Pump (CP-505) - 2 Items	$8,400

Type: Centrifugal, AVS medium
Capacity: 600 GPM
Material: 304 SS
Head: 50' TDH
Drive: 25 HP
Temperature: 100°F

Distillate Pump (CP-506) - 2 Items	$8,400

Type: Centrifugal, AVS medium
Capacity: 690 GPM
Material: CS
Head: 100' TDH
Drive: 25 HP
Temperature: 180°F

Subtotal Section 500	$3,699,700
Undefined Miscellaneous	31,300
Total Section 500	$3,731,000

SECTION 600 - FERMENTATION

	Cost
Concentrated Sugar Receiver (VT-608)	$25,500

Type: Cylinder
Capacity: 7,000 gallons
Material: 304 SS
Temperature: 100°F
Pressure: 0.05 psig
Holdup: 10 minutes

Blower (FN-601)	$37,500

Type: ROT blower
Capacity: 3,040 CFM
Material: CS
Drive: 75 HP
Pressure: 5 psig

Air Blower (FN-602)	$10,000

Type: ROT blower
Capacity: 100 CFM
Material: CS
Drive: 10 HP
Pressure: 5 psig

Filter (F-601)	$2,000

Area: 50 square feet
Type: Cartridge
Material: CS

Filter (F-602)	$2,000

Area: 50 square feet
Type: Cartridge
Material: CS

Fermentation Feed Pump (CP-613) - 2 Items	$8,400

Type: Centrifugal, AVS medium
Capacity: 600 GPM
Material: 304 SS
Head: 100' TDH
Drive: 75 HP
Temperature: 100°F

Sugar Cooler (HE-607) $27,300

Area: 805 square feet
Type: Spiral plate
Duty: 3.5 MM Btu/hr
Material: CS/304 SS
Temperature: 80/100°F
Pressure: 15/15 psig

Fermentation Tanks (VT-601-606) - 6 Items $614,400

Type: Cone roof
Capacity: 150,000 GPM
Material: 304 SS
Temperature: 86°F
Pressure: 0.05 psig

Recirculation Pumps (CP-601-606) - 12 Items $50,400

Type: Centrifugal, AVS medium
Capacity: 600 GPM
Material: 304 SS
Head: 100' TDH
Drive: 25 HP
Temperature: 86°F

Discharge Pumps (CP-607-612) - 12 Items $50,400

Type: Centrifugal, AVS medium
Capacity: 600 GPM
Material: 304 SS
Head: 100' TDH
Drive: 25 HP
Temperature: 86°F

Coolers (HE-601-606) - 6 Items $146,400

Area: 710 square feet
Type: Spiral plate
Duty: 2.2 MM Btu/hr
Material: CS/304 SS
Temperature: 80/86°F
Pressure: 15/15 psig

Scrubber (TW-601) $20,000

Type: Packed
Size: 4' diameter x 10' t/t
Material: 304 SS

Scrubber Return Pump (CP-613) - 2 Items $5,400

Type: Centrifugal, AVS medium
Capacity: 50 GPM
Material: 304 SS
Head: 100' TDH
Drive: 2 HP
Temperature: 85°F

Yeast Centrifuge (CT-601) - 2 Items $229,600

Type: Disk
Size: 30" diameter
Material: 304 SS
Drive: 200 HP
Overflow: 300 GPM

Overflow Pump (CP-616) - 4 Items $16,000

Type: Centrifugal, AVS medium
Capacity: 300 GPM
Material: 304 SS
Head: 100' TDH
Drive: 15 HP
Temperature: 36°F

Underflow Pump (CP-615) - 4 Items $10,400

Type: Centrifugal, AVS medium
Capacity: 25 GPM
Material: 304 SS
Head: 100' TDH
Drive: 3 HP
Temperature: 86°F

Yeast Hold Tank (AOT-601) $68,200

Type: Mixer
Volume: 10,000 gallons
Material: 304 SS
Drive: 40 HP
Temperature: 86°F
Pressure: 0.05 psig
Holdup: 4 hours

Yeast Pump (CP-617) - 2 Items $5,200

Type: Centrifugal, AVS medium
Capacity: 40 GPM
Material: 304 SS
Head: 100' TDH
Drive: 3 HP
Temperature: 86°F

Dewatering Press (F-603)	$45,000

Type: Fibercone Press
Capacity: 3 GPM
Material: CS
Drive: 1 HP

Discharge Pump (CP-618) - 2 Items	$4,000

Type: Centrifugal, AVS medium
Capacity: 5 GPM
Material: 304 SS
Head: 50' TDH
Drive: 1 HP
Temperature: 36°F

Conveyor (CO-601)	$10,000

Type: Screw
Capacity: 0.5 tons/hr
Material: CS
Drive: 7.5 HP
Length: 100'

Beer Well (VT-607) - 2 Items	$204,800

Type: Cone roof
Capacity: 150,000 gallons
Material: 304 SS
Temperature: 86°F
Pressure: 0.05 psig
Holdup: 8 hours

Purification Feed Pump (CP-619) - 2 Items	$8,400

Type: Centrifugal, AVS medium
Capacity: 600 GPM
Material: 304 SS
Head: 180' TDH
Drive: 40 HP
Temperature: 86°F

Subtotal Section 600	$1,601,300
Undefined Miscellaneous	112,700
Total Section 600	$1,714,000

SECTION 700 - CARBON DIOXIDE RECOVERY

	Cost
Package Unit	$5,800,000
Capacity: 300 tons/day Drives: 2,100 HP	
Total Section 700	$5,800,000

SECTION 800 - ETHANOL PURIFICATION

	Cost
Beer Column (TW-801)	$324,000

Type: Distillation Column
Size: 9.5' dia. x 135' t/t
Material: CS shell/304 SS trays
Temperature: 280°F
Pressure: 40 psig
Trays: 30 sieve
30 disc and donut

First Stage Beer Preheater (HE-801) — $29,800

Area: 2,400 square feet
Type: Fixed T-S
Duty: 16.1 MM Btu/hr
Material: CS/CS
Temperature: 160/138°F
Pressure: 15/30 psig

Second Stage Beer Preheater (HE-802) — $16,600

Area: 1,100 square feet
Type: Fixed T-S
Duty: 10.2 MM Btu/hr
Material: CS/CS
Temperature: 230/185°F
Pressure: 30/30 psig

Trim Condenser (HE-805) — $23,000

Area: 1,700 square feet
Type: Fixed T-S
Duty: 34.2 MM Btu/hr
Material: CS/CS
Temperature: 100/104°F
Pressure: 30/150 psig
Service: Start-up

Vent Condenser (HE-806) — $4,400

Area: 100 square feet
Type: U-tube
Duty: 0.5 MM Btu/hr
Material: CS

Beer Column Bottoms Pump (CP-801) - 2 Items — $9,200

Type: Centrifugal, AVS medium
Capacity: 1,700 GPM

Material: 304 SS
Head: 50' TDH
Drive: 25 HP
Temperature: 280°F

Reflux Pump (CP-802) - 2 Items — $7,800

Type: Centrifugal, AVS medium
Capacity: 300 GPM
Material: 304 SS
Head: 150' TDH
Drive: 20 HP
Temperature: 235°F

Preheater (HE-813) — $23,000

Area: 1,700 square feet
Type: Fixed T-S
Duty: 9.2 MM Btu/hr
Material: CS/CS
Temperature: 200/300°F
Pressure: 30/30 psig

Final Beer Preheater (HE-803) — $10,000

Area: 500 square feet
Type: Fixed T-S
Duty: 7.8 MM Btu/hr
Material: CS/CS
Temperature: 250/300°F
Pressure: 30/50 psig

Beer Column Reboiler (HE-804) — $18,800

Area: 1,300 square feet
Type: Fixed T-S
Duty: 52.4 MM Btu/hr
Material: CS/CS
Temperature: 386/280°F
Pressure: 200/30 psig

Reflux Drum (VT-801) — $12,600

Type: Vertical
Capacity: 1,500 GPM
Size: 5' diameter x 10' t/t
Material: 304 SS
Temperature: 230°F
Pressure: 35 psig

Dehydration Column (TW-802) $247,000

Type: Distillation column
Size: 5' diameter x 115' t/t
Material: CS
Temperature: 200°F
Pressure: 0.05 psig
Trays: 50 sieve

Dehydration Column Reboiler (HE-807) $120,000

Area: 10,000 square feet
Type: Fixed T-S
Duty: 29.4 MM Btu/hr
Material: CS/CS
Temperature: 199/230°F
Pressure: 0.05/50 psig

Dehydration System Overhead Cooler (HE-808) $38,000

Area: 3,200 square feet
Type: Fixed T-S
Duty: 18.4 MM Btu/hr
Material: CS/CS
Temperature: 176/100°F
Pressure: 15/150 psig

Decanter (HT-801) $19,700

Type: Horizontal drum
Capacity: 8,800 gallons
Material: CS
Temperature: 175°F
Pressure: 0.05 psig

Dehydration Column Reflux Pump (CP-804) - 2 Items $7,800

Type: Centrifugal, AVS medium
Capacity: 240 GPM
Material: 304 SS
Head: 50' TDH
Drive: 5 HP
Temperature: 150°F

Entrainer/Stripper Column Feed Pump (CP-805) - 2 Items $5,000

Type: Centrifugal, AVS medium
Capacity: 20 GPM
Material: 304 SS
Head: 50' TDH
Drive: 2 HP
Temperature: 150°F

Entrainer/Stripper Column (TW-803) $80,500

Type: Distillation column
Size: 3' diameter x 60' t/t
Material: CS
Temperature: 205°F
Pressure: 0.05 psig
Trays: 30 sieve

Entrainer Stripper Reboiler (HE-809) $24,400

Area: 1,900 square feet
Type: Fixed T-S
Duty: 4.8 MM Btu/hr
Material: CS/CS
Temperature: 230/205°F
Pressure: 50/5 psig

Entrainer Column Bottoms Pump (CP-806) - 2 Items $6,400

Type: Centrifugal, AVS medium
Capacity: 60 GPM
Material: 304 SS
Head: 100' TDH
Drive: 2 HP
Temperature: 205°F

Fusel Oil Washer/Decanter (DDT-801) $14,000

Type: Vertical
Size: Top - 3' diameter x 2' t/t
Bottom - 2' diameter x 10' t/t
Material: CS
Temperature: 100°F
Pressure: 0.05 psig

Fusel Oil Hold Tank (VT-802) $4,300

Type: Vertical
Capacity: 100 gallons
Material: CS
Temperature: 105°F
Pressure: 0.05 psig

Fusel Oil Cooler (HE-810) $4,400

Area: 100 square feet
Type: U-tube
Material: CS

Wash Water Pump (CP-809) - 2 Items $3,000

Type: Centrifugal, AVS medium
Capacity: 2 GPM

Material: CS
Head: 150' TDH
Drive: 0.5 HP
Temperature: 100°F

Fusel Oil Pump (CP-808) - 2 Items	$3,000

Type: Centrifugal, AVS medium
Capacity: 2 GPM
Material: CS
Head: 100' TDH
Drive: 0.5 HP
Temperature: 100°F

Waste Water Cooler (HE-811)	$4,400

Area: 100 square feet
Type: U-tube
Duty: 0.1 MM Btu/hr
Material: CS
Temperature: 205°F
Pressure: 50 psig

Alcohol Cooler (HE-812)	$10,200

Area: 300 square feet
Type: U-tube
Duty: 0.6 MM Btu/hr
Material: CS
Temperature: 200°F
Pressure: 5 psig

Dehydration Column Bottoms Pump (CP-803) - 2 Items	$5,400

Type: Centrifugal, AVS medium
Capacity: 60 GPM
Material: CS
Head: 100' TDH
Drive: 2 HP
Temperature: 200°F

Subtotal Section 800	$1,077,200
Undefined Miscellaneous	75,800
Total Section 800	$1,153,000

SECTION 900 - FURFURAL PRODUCTION

	Cost
Stillage Receiver (VT-901)	$78,700
Type: Cone roof Capacity: 270,000 gallons Material: CS Temperature: 280°F Pressure: 75 psig Holdup: 8 hours	
Furfural Reactor Feed Pump (CP-901) - 2 Items	$11,200
Type: Centrifugal, AVS medium Capacity: 550 GPM Material: CS Head: 200' TDH Drive: 40 HP Temperature: 280°F	
Furfural Reactor (R-901)	$580,000
Type: Tower w/baffles Capacity: 1,000 gallons Material: Zirconium clad Temperature: 428°F Pressure: 700 psig	
First Flash Vessel (VT-902)	$42,000
Type: Cylinder Capacity: 1,000 gallons Material: Zirconium Temperature: 250°F Pressure: 30 psig	
Second Flash Vessel (VT-903)	$28,200
Type: Cylinder Capacity: 700 gallons Material: 317 SS Temperature: 212°F Pressure: 15 psig	
Neutralization Tank (AOT-902)	$49,700
Type: Mixer Capacity: 5,000 gallons Material: 317 SS	

Drive: 25 HP
Temperature: 212^oF
Pressure: 15 psig

Lime Pump (CP-902) - 2 Items $4,000

Type: Centrifugal, AVS medium
Capacity: 10 GPM
Material: CS
Head: 100' TDH
Drive: 1 HP
Temperature: 80^oF

Lime Tank (AOT-901) $29,500

Type: Mixer
Volume: 3,500 gallons
Material: CS
Drive: 20 HP
Temperature: 80^oF
Pressure: 0.05 psig
Holdup: 8 hours

Thickener Feed Pump (CP-903) - 2 Items $7,600

Type: Centrifugal, AVS medium
Capacity: 450 GPM
Material: CS
Head: 100' TDH
Drive: 20 HP
Temperature: 212^oF

Thickener (T-901) $34,300

Type: Gravity with rake
Capacity: 450 GPM
Material: CS
Drive: 5 HP
Diameter: 24'

$CaSO_4$ Centrifuge (CT-901) $56,400

Type: Solid bowl
Size: 10" diameter
Material: CS
Drive: 25 HP
Overflow: 60 GPM

Overflow Pump (CP-904) - 2 Items $5,400

Type: Centrifugal, AVS medium
Capacity: 60 GPM
Material: CS

Head: 100' TDH
Drive: 2 HP
Temperature: 212°F

Intermediate Furfural Receiver (VT-904) $32,000

Type: Cone roof
Capacity: 60,000 gallons
Material: CS
Temperature: 212°F
Pressure: 0.05 psig
Holdup: 2 hours

Azeotrope Feed Pump (CP-905) -2 Items $7,600

Type: Centrifugal, AVS medium
Capacity: 500 GPM
Material: CS
Head: 100' TDH
Drive: 20 HP
Temperature: 212°F

Azeotrope Reboiler (HE-901) $96,400

Area: 3,200 square feet
Type: Fixed T-S
Duty: 50.8 MM Btu/hr
Material: CS/304 SS
Temperature: 248/220°F
Pressure: 30/30 psig

Azeotrope Column (TW-901) $234,000

Type: Distillation column
Size: 8' dia. x 140' t/t
Material: Shell - CS
Trays - 304 SS
Pressure: 15 psig
Trays: 70 sieve

Azeotrope Condenser (HE-908) $31,000

Area: 790 square feet
Type: Fixed T-S
Duty: 48.7 MM Btu/hr
Material: CS/304 SS
Temperature: 122/208°F
Pressure: 15/15 psig

BFW Preheater (HE-903) $19,500

Area: 460 square feet
Type: Fixed T-S
Duty: 16.6 MM Btu/hr

Material: CS/304 SS
Temperature: 180/208°F
Pressure: 15/15 psig

<u>Decanter (HT-901)</u> $14,900

Type: Horizontal
Capacity: 6,000 gallons
Material: CS
Temperature: 208°F
Pressure: 0.05 psig
Holdup: 30 minutes

<u>Lights Column Feed Pump (CP-910) - 2 Items</u> $7,000

Type: Centrifugal, AVS medium
Capacity: 160 GPM
Material: CS
Head: 100' TDH
Drive: 7.5 HP
Temperature: 208°F

<u>Dehydration Feed Pump (CP-907) - 2 Items</u> $4,400

Type: Centrifugal, AVS medium
Capacity: 15 GPM
Material: CS
Head: 100' TDH
Drive: 1 HP
Temperature: 208°F

<u>Lights Column (TW-902)</u> $61,000

Type: Distillation column
Size: 3' dia. x 30' t/t
Material: Shell - CS
Trays - 304 SS
Pressure: 15 psig
Trays: 15 sieve

<u>Reflux Pump (CP-912) - 2 Items</u> $4,000

Type: Centrifugal, AVS medium
Capacity: 5 GPM
Material: CS
Head: 100' TDH
Drive: 0.5 HP
Temperature: 176°F

<u>Lights Column Condenser (HE-905)</u> $9,700

Area: 200 square feet
Type: Fixed T-S

Duty: 9.0 MM Btu/hr
Material: CS/304 SS
Temperature: 120/176°F
Pressure: 15/15 psig

Reflux Drum (HT-902) $1,000

Type: Horizontal
Capacity: 100 gallons
Material: CS
Temperature: 176°F
Pressure: 0.05 psig

Lights Cooler (HE-909) $1,000

Area: 2 square feet
Type: Fixed T-S
Duty: 0.02 MM Btu/hr
Material: CS/CS
Temperature: 120/176°F
Pressure: 15/15 psig

Lights Column Reboiler (HE-904) $19,600

Area: 410 square feet
Type: Fixed T-S
Duty: 9.0 MM Btu/hr
Material: CS/304 SS
Temperature: 248/212°F
Pressure: 30/15 psig

Lights Bottoms Pump (CP-911) - 2 Items $7,200

Type: Centrifugal, AVS medium
Capacity: 180 GPM
Material: CS
Head: 100' TDH
Drive: 7.5 HP
Temperature: 212°F

Azeotrope Bottoms Pump (CP-906) - 2 Items $8,800

Type: Centrifugal, AVS medium
Capacity: 725 GPM
Material: CS
Head: 100' TDH
Drive: 25 HP
Temperature: 220°F

Stillage Pump (CP-913) - 2 Items $8,800

Type: Centrifugal, AVS medium
Capacity: 650 GPM

Material: CS
Head: 150' TDH
Drive: 25 HP
Temperature: 220°F

Dehydration Column (TW-803) $5,000

Type: Distillation column
Size: 1' dia. x 20' t/t
Material: Shell - CS
Trays - 304 SS
Pressure: 15 psig
Trays: 10 sieve

Dehydration Condenser (HE-907) $6,600

Area: 20 square feet
Type: Fixed T-S
Duty: 1.2 MM Btu/hr
Material: CS/304 SS
Temperature: 120/208°F
Pressure: 15/15 psig

Reflux Drum (HT-903) $1,000

Type: Horizontal
Capacity: 100 gallons
Material: CS
Temperature: 208°F
Pressure: 0.05 psig

Reflux Pump (CP-909) - 2 Items $4,000

Type: Centrifugal, AVS medium
Capacity: 5 GPM
Material: CS
Head: 50' TDH
Drive: 0.5 HP
Temperature: 208°F

Dehydration Reboiler (HE-906) $7,000

Area: 30 square feet
Type: Fixed T-S
Duty: 1.2 MM Btu/hr
Material: CS/304 SS
Temperature: 400/324°F
Pressure: 250/15 psig

Dehydration Bottoms Pump (CP-908) - 2 Items $4,600

Type: Centrifugal, AVS medium
Capacity: 25 GPM

Material: CS
Head: 100' TDH
Drive: 2 HP
Temperature: 324°F

Furfural Cooler (HE-902) $8,000

Area: 120 square feet
Type: Fixed T-S
Duty: 1.0 MM Btu/hr
Material: CS/304 SS
Temperature: 120/324°F
Pressure: 15/15 psig

Subtotal Section 900	$1,531,000
Undefined Miscellaneous	107,000
Total Section 900	$1,638,000

SECTION 1000 - HEAT GENERATION

	Cost
Multi-Effect Evaporator (E-1001)	$2,100,000
Type: Six effect-forced circulation Duty: Heating - 44.8 MM Btu/hr Cooling - 67.8 MM Btu/hr Capacity: 328,000 lb/hr distillate Material: CS Drive: 1,100 HP	
Distillate Pump (CP-1006) - 2 Items	$10,000
Type: Centrifugal, AVS medium Capacity: 660 GPM Material: CS Head: 150' TDH Drive: 25 HP Temperature: 169°F	
Wood Boiler Package	-
Part of offsite facilities	
Total Section 1000	2,110,000

SECTION 1100 - WASTE TREATMENT

	Cost
Condensate Treatment System (WTS-1101)	$2,060,000

Package Unit
Type: Aerobic/anaerobic carrousel system
Capacity: 1,450 GPM
Size: Aerobic basins - 2-177' diameter
Anaerobic clarifiers - 2-65' diameter
Gravity thickener - 1-30' diameter
Drive: 600 HP

Gypstack (WTS-1102)	$500,000

Area: 1 acre expandable
Type: Dyked/lined storage
Drives: 60 HP

Total Section 1100	$2,560,000

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15. Supplementary Notes

Technical Monitor: John Wright

16. Abstract (Limit: 200 words) This report presents an economic feasibility study of an enzymatic-hydrolysis-based ethanol plant with prehydrolysis treatment. Dilute-acid prehydrolysis had been determined to be one of the most economic pretreatments for enzyme hydrolysis. A Midwest site was selected for the ethanol facility; screening parameters were ethanol consumption patterns, gasoline demand, tax laws, and wood resources. The facility is designed to produce 25 million gallons per year of fuel-grade ethanol from mixed hardwood chips, with furfural and carbon dioxide as major by-products. It was determined that an enzymatic hydrolysis based ethanol plant using prehydrolysis pretreatment is a viable process concept.

17. Document Analysis

a. Descriptors Feasibility studies, ethanol, hydrolysis, enzymatic hydrolysis, furfural

b. Identifiers/Open-Ended Terms

c. UC Categories

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